Donlin Gold
Advanced Water Treatment

Advanced Water Treatment Options Report

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1. Introduction

The Donlin Gold project has progressed through Feasibility level design and an Environmental Impact Statement is currently being prepared by the U.S. Army Corps of Engineers as required under the National Environmental Policy Act (NEPA).

The site anticipates a positive water balance and will require a water treatment plant (WTP) to allow inventories of site water to be managed and released in compliance with discharge objectives. There are four sources of water at site that will require treatment prior to discharge in order to meet discharge quality standards that are defined based on the most stringent criteria for the State of Alaska:

- Perimeter and In-Pit Dewatering Wells;
- Upper Contact Water Dam (Upper CWD);
- Seepage Recovery System (SRS); and
- Tailings Storage Facility (TSF).

The WTP was originally designed based on a conceptualized high density sludge (HDS) process to only treat water from the dewatering wells. Of the four sources, water from the dewatering wells generally requires the lowest level of treatment to meet the discharge objectives. However, the project has developed such that treatment and discharge of SRS and Upper CWD water are desirable, both waters having higher predicted metal and salt concentrations than the dewatering wells. The HDS process is limited in its capacity for treatment of SRS and Upper CWD water due to the degree to which it is able to remove certain parameters to meet discharge objectives, particularly for total dissolved solids (TDS). Additionally, the HDS process is not designed to treat water from the TSF.

Donlin Gold LLC (Donlin Gold) would like to evaluate potential treatment options for the other sources of water. This would provide a tool for more flexible site water management, including the ability to draw down storage inventory in the TSF. This report will address the different treatment options and technologies selection. A description of the proposed process will follow along with a preliminary flow diagram, equipment arrangement and discussion of anticipated performance.

2. List of Abbreviations and Acronyms

The following list of abbreviations and acronyms will be used in this report:

AMD	Acid Mine Drainage
AWT	Advanced Water Treatment
CIP	Clean-In-Place
CWD	Contact Water Dam
EC	Electrocoagulation
FRP	Fiber Reinforced Plastic
HDS	High Density Sludge
HRC	High Rate Clarifier
HRT	Hydraulic Retention Time
IX	Ion Exchange
MF	Microfiltration
MMF	Multimedia Filtration
NF	Nanofiltration
PP	Polypropylene
RO	Reverse Osmosis
SAC	Strong Acid Cation
SBA	Strong Base Anion
SDI	Silt Density Index
SRS	Seepage Recovery System
SS	Stainless Steel
TDS	Total Dissolved Solids
TSF	Tailings Storage Facility
TSS	Total Suspended Solids
UF	Ultrafiltration
WTP	Water Treatment Plant

3. Objectives

The main objective for the advanced water treatment study is to determine an advanced water treatment technology that can meet all discharge objectives for discharge of excess water.

Other objectives are to:

- Control water accumulation in the TSF to maintain an inventory of about 10 Mm^3; and
- Provide a tool to enhance operational flexibility.

4. Treatment Process

4.1 Design Basis

The WTP is expected to operate seasonally, but will have the ability to operate and discharge year round if required. Treatment will primarily be required in the summer period (April to October) when there is expected to be an excess of water in the Lower CWD and Upper CWD for process use. By contrast, in the winter period (November to March), the CWD pond volumes are typically low and the dewatering well water and SRS water will be required as a source of make-up for the process. The WTP will operate at a minimum throughput, or be dormant through the winter [12].

Design basis water qualities and flows are based on the Preliminary Design Basis [11].

4.1.1 Flowrates

4.1.1.1 Maximum Flows

Table 4-1 presents the maximum flows for the four individual water sources during the life of mine (Refer to [11]). The maximum flow of TSF reclaim water to the WTP has been set at 71 m^3/h. This limit is based on the mercury concentration in the reclaim water and expected removal efficiency in the WTP (see Section 4.4.4.3). Treatment of TSF water at a greater rate may be possible depending on actual water quality. The flow of TSF reclaim water that may be treated and discharged is further constrained such that the annual volume does not exceed the anticipated excess precipitation falling on the TSF. This constraint limits TSF treatment to 12 m^3/h in year -1 and to 50 m^3/h in years 1 to 4.

Table 4-1: Maximum Flows to WTP During the Life of Mine

Year	Maximum Flows to WTP (m^3/h)				
	Wells	SRS	Upper CWD	TSF	Total
-2	334	161	250	0	745
-1	329	161	250	12	752
1	269	161	250	50	730
2	316	161	250	50	777
3	412	235	250	50	947
4	355	235	250	50	890
5	343	235	250	71	899
6	323	235	250	71	879
7	315	235	250	71	871
8	392	235	250	71	948
9	369	235	250	71	925
10	364	200	250	71	885
11	360	200	250	71	881
12	540	200	250	71	1061
13	467	200	250	71	988

Year	Maximum Flows to WTP (m^3/h)				
	Wells	SRS	Upper CWD	TSF	Total
14	419	200	250	71	940
15	396	169	250	71	886
16	419	169	250	71	909
17	397	169	250	71	887
18	400	169	250	71	890
19	394	169	250	71	884
20	212	153	250	71	686
21	239	153	250	71	713
22	253	153	250	71	727
23	261	153	250	71	735
24	265	153	250	71	739
25	28	134	250	71	483

Notes:
- Highlighted values have been assumed. BGC [3] has predicted flows from the SRS only in years 2, 5, 10,15, 20, and 25. Assumed values are set equal to the highest flow predicted by BGC in year s bracketing the unknown flow.
- Groundwater pumping in the pit area will end in year 25.

From Table 4-1, a maximum total flow to the WTP of 1061 m^3/h will occur in year 12. This value will be used as the design WTP capacity. If required, the capacity of the plant could be expanded to 1500 m^3/d.

4.1.1.2 *Average Flows*

Table 4-2 shows average flows for the four individual water sources during the life of mine. Average flows shown are for the summer period (April to October).

Table 4-2: Average Flows to WTP During the Life of Mine (Summer Period)

Average Flows to WTP (m^3/h)				
Wells	SRS	Upper CWD	TSF	Total
318	165	156	30	669

An average flow during the life of mine of 669 m^3/h will be used as basis for design.

4.1.2 *Influent Water Quality*

An influent water quality has been calculated for the treatment process based on blending of the four individual sources of water at their respective average flow from Table 4-2. A worst-case water quality has been estimated from the blend of the individual sources at their 95th percentile water quality. The blended water quality has also been estimated using 50th percentile water quality values from each individual source. The TSF reclaim and pore water qualities used in these calculations represent the steady-state condition [7]. Mercury concentrations used for the TSF reclaim and pore water quality in the calculation of blended feed have been adjusted to apply an assumed maximum concentration of 0.01 mg/L based on experience with use of UNR reagent at Pueblo Viejo [17].

CWD water quality used in calculating the blended water quality is representative of the Upper CWD. Upper CWD water quality has been determined based on the mixture of pumped water from the Lower CWD with run-off water to the Upper CWD (refer to [11]).

The anticipated WTP feed water quality and water quality for each individual source are presented in Table 4-3. It should be noted that the methodology used to estimate the blended feed assumes that 95th percentile concentrations occur simultaneously in the dewatering wells, Upper CWD and SRS. This is not typically the case, so the design basis water quality given in Table 4-3 is expected to be conservative. Lower CWD water may also be treated and discharged at a higher rate if the water quality is better than these 95th percentile predictions, for example during rainy periods when seepage is a lower fraction of the flow. Feed water quality values exceeding the anticipated discharge objectives (refer to Section 4.1.3) are highlighted in green.

The blended WTP feed prediction resulting from use of the 95th percentile concentrations for the individual sources will be used for WTP design and performance evaluation. An exception has been made for mercury. In the case of mercury, WTP performance and limitations have been evaluated assuming that there will be restrictions on treatment of water from the SRS after year 18 during low flow periods, typically in late winter (refer to Appendix A). The maximum mercury concentration considered for feed to the WTP is 0.000288 mg/L.

Table 4-3: WTP Influent Water Quality

Parameter	Unit	Wells		SRS		Upper CWD		TSF	WTP Blended Feed	
		50th Percentile	95th Percentile	50th Percentile	95th Percentile	50th Percentile	95th Percentile	Steady-State	50th Percentile	95th Percentile
Aluminum	mg/L	0.081	2.67	0.019	0.043	0.073	0.121	0.013	0.061	1.309
Ammonia	mg/L	0.315	1.1	3.1	3.2	0.541	0.551	29	2.341	2.741
Antimony	mg/L	<0.0005	<0.001	0.12	0.12	0.423	0.483	0.022	0.135	0.149
Arsenic	mg/L	0.1	2.13	1.61	1.61	2.444	2.871	3.3	1.164	2.228
Barium	mg/L	0.151	1.45	0.06	0.09	0.122	0.176	0.011	0.115	0.753
Beryllium	mg/L	<0.0004	<0.0004	0.0004	0.002	0.00048	0.00053	<0.00006	0.00022	0.00063
Boron	mg/L	0.03	0.19	0.085	0.11	0.110	0.114	0.594	0.073	0.081
Cadmium	mg/L	<0.00025	<0.0005	0.0003	0.0052	0.0008	0.0008	0.00073	0.00041	0.00059
Calcium	mg/L	33.35	62.79	122	128	111	123	610	116.6	135.0
Chloride	mg/L	1.21	3.4	3.2	3.6	1.0	1.3	26	2.7	3.9
Chromium, total	mg/L	<0.001	0.0043	0.002	0.003	0.013	0.014	0.012	0.005	0.005
Cobalt	mg/L	<0.002	<0.004	0.0022	0.0031	0.032	0.032	0.019	0.010	0.011
Copper	mg/L	0.0015	0.0098	0.0024	0.0028	0.005	0.005	0.018	0.003	0.007
Fluoride	mg/L	0.4	0.52	0.303	0.303	0.153	0.159	2	0.390	0.449
Iron	mg/L	1.76	5.5	10.6	10.7	0.30	0.47	0.0044	3.6	5.4
Lead	mg/L	<0.0001	<0.0027	0.0004	0.0005	0.0424	0.0485	0.003	0.0102	0.0128
Lithium	mg/L	0.011	0.014	0.005	0.005	0.062	0.072	<0.006	0.030	0.034
Magnesium	mg/L	10.35	26.33	110	112	14.4	18.1	1734	113.2	122.1
Manganese	mg/L	0.22	1.41	0.23	0.27	0.95	1.26	2	0.5	1.1
Mercury[1]	mg/L	0.0000022	0.000014	0.00107	0.00108	0.000099	0.000108	0.010	0.00074	0.00075
Molybdenum	mg/L	<0.005	<0.01	0.029	0.034	0.101	0.123	0.23	0.044	0.052
Nickel	mg/L	<0.001	0.005	0.0075	0.0084	0.187	0.189	0.062	0.049	0.051
pH	--	7.34	8.8	7.2	7.8	7.93	8.06	7.7	7.46	8.33
Potassium	mg/L	1.18	3.93	13.2	15.4	13.0	16.7	120	12.2	14.9
Selenium	mg/L	<0.0025	<0.005	0.0067	0.0067	0.125	0.136	0.042	0.034	0.038
Silica	mg/L	5.49	10.58	N/A	N/A	16.0	16.9	7	7.89	10.52

Parameter	Unit	Wells		SRS		Upper CWD		TSF	WTP Blended Feed	
		50th Percentile	95th Percentile	50th Percentile	95th Percentile	50th Percentile	95th Percentile	Steady-State	50th Percentile	95th Percentile
Sodium	mg/L	38.6	132.2	120.7	122.1	12.3	21.7	1100	100.3	147.3
Strontium	mg/L	0.35	1.01	N/A	N/A	1.65	2.05	7.9	1.08	1.48
Sulphate	mg/L	9.1	81.9	471	474	279	333	8612	572	620
TDS	mg/L	178	545	905	919	519	614	11550	966	1166
Thallium	mg/L	<0.0005	<0.0005	0.0005	0.001	0.001	0.001	0.00041	0.0006	0.0007
TSS	mg/L	6.29	61.8	N/A	N/A	N/A	N/A	21	3.9	30.3
Vanadium	mg/L	<0.01	<0.02	0.005	0.02	0.013	0.018	0.0048	0.0092	0.0188
WAD Cyanide[2]	mg/L	<0.0025	<0.0025	0.016	0.018	0.005	0.005	0.13	0.015	0.017
WAD Cyanide[3]	mg/L	<0.0025	<0.0025	0.030	0.032	0.005	0.005	0.26	0.021	0.023
Zinc	mg/L	<0.0025	0.03	0.006	0.008	0.267	0.281	0.033	0.066	0.083

Notes: [1] Mercury concentrations are after year 18 during late winter with minimal SRS underdrain flow. The maximum WTP feed concentration considered for design is 0.000288 mg/L.
[2] Summer values. 50% natural degradation assumed in the TSF.
[3] Winter values. Assumes no natural degradation in the TSF.
- N/A: Not available
- Parameters highlighted in green exceed the most stringent water quality objectives

4.1.3 *Treatment Objectives*

Crooked Creek is the proposed discharge location for treated water from the WTP. Crooked Creek water is classified for all uses; therefore, the most stringent Alaskan water quality standard (WQS) is to be used to define the treated water discharge objectives at end-of-pipe.

Table 4-4 provides the discharge objectives for the Donlin Gold site for each water parameter. The provided values are the water quality objectives from which maximum daily limit (MDL) and average monthly limit (AML) values would be derived for permitting.

Table 4-4: Donlin Gold Water Quality Treatment Objectives

Parameter	Donlin Gold Treatment Objectives	
	Treatment Objective (mg/L)	Basis for Objective
Alkalinity	>20	AQ (except where natural conditions are lower)
Aluminum[1]	0.087	AQ
Ammonia[2]	2.16	AQ
Antimony	0.006	HH
Arsenic	0.01	HH
Barium	2	HH
Beryllium	0.004	DW
BOD	n/a	Consideration to be given to DO in receiving body
Boron	0.75	Irrig
Cadmium	0.00025	AQ
Calcium	n/a	TDS standard applies at high concentrations
Chloride	230	AQ
Chromium, total	0.1	HH
Chromium III	0.074	AQ
Chromium VI	0.011	AQ
Cobalt	0.05	Irrig
Copper	0.0089	AQ
Fluoride	1	Irrig
Iron	1	AQ
Lead	0.0032	AQ
Lithium	2.5	Irrig
Magnesium	n/a	TDS standard applies at high concentrations
Manganese	0.05	HH
Mercury	0.000012	AQ
Molybdenum	0.01	Irrig
Nickel	0.052	AQ
Nitrate	10	HH
pH	6.5 - 8.5	AQ
Phosphate	n/a	Mass loading may need to be minimized
Potassium	n/a	TDS standard applies at high concentrations
Selenium	0.005	AQ
Silica	n/a	TDS standard applies at high concentrations
Sodium	n/a	TDS standard applies at high concentrations
Strontium	n/a	TDS standard applies at high concentrations
Sulphate	250	DW
TDS	500	DW
Temperature	13 °C	AQ
Thallium	0.0017	HH
TSS	20	Effluent Limitation Guidelines Subpart J
Vanadium	0.1	Irrig
WAD Cyanide	0.0052	AQ
Zinc	0.119	AQ

Notes: [1]At monitoring station CCBW the 15th percentile hardness is 99.64 mg/L and the 15th percentile pH is 6.9. The 15th percentiles were derived from data collected from 2005 through 2014. Where the pH is greater than or equal to 7.0 and the hardness is greater than or equal to 50 mg/L as $CaCO_3$, the chronic aluminum standard is equal to the acute aluminum standard, 0.75 mg/L (Alaska Water Quality Criteria Manual, 2008).

[2]Based on a pH of 8.08 and a temperature of 6.13° C, which represents the 85th percentile values at Station CCBW.

Hardness-based standards for metals were derived using a hardness of 99.64 mg/L, which represents the 15th percentile measured hardness at Station CCBW, Crooked Creek.

AQ: Aquatic life water quality standard for chronic conditions.

DW: Drinking water quality standard.

HH: Human health water quality standard.

Irrig: Irrigation water quality standard.

4.2 Evaluation and Selection of Technologies

Table 4-5 presents the technologies evaluated during the Advanced Water Treatment (AWT) assessment. Assessed technologies have been categorized as regards to their applicability for pre-treatment and/or polishing of the Donlin Gold influent. Pre-treatment is defined as all steps involved in reducing influent feed concentrations to levels suitable for polishing. Polishing is defined as any step intended to produce treated effluent complying with the discharge objectives. An AWT process may consist of multiple pre-treatment steps together with one or more polishing technologies.

Table 4-5: Assessed Technologies for Advanced Water Treatment

Technology	Pre-treatment	Polishing
High Density Sludge	X	
Iron Co-precipitation / Adsorption	X	
Ferrous Hydroxide Reduction	X	
Electrocoagulation	X	
Multimedia Filtration	X	
Manganese Greensand Filtration	X	
Microfiltration / Ultrafiltration	X	
Ion Exchange	X	X
Nanofiltration		X
Reverse Osmosis		X
Evaporation and Crystallization		X

4.2.1 Pre-Treatment Technology Options

4.2.1.1 High Density Sludge

The high density sludge (HDS) process carries out an effective removal of base metals in a chemically stable form, primarily as the result of the formation of co-precipitates with iron on the surfaces of the recycled sludge particles. The chemical stability of the precipitates is favourably influenced by a high iron to total metals ratio in the treatment plant feed. The HDS process produces a sludge with a high percent of solids. High iron, alum and sulphate concentrations are required to form a dense sludge.

HDS is a proven technology which has been broadly applied in the mining industry. Lime and a ferric salt, if there is insufficient iron in the feed, are required.

The HDS process is normally run at a pH between 9.0 and 9.5, as most metals will precipitate at or below this pH. However, some species, such as cadmium, manganese and mercury, may need a higher pH (9.5 – 10) for effective removal.

There is a good removal of most of the species present in mining wastewater, except for ammonia, cyanide, molybdenum, selenium, sulphate and total dissolved solids (TDS).

Other parameters, like aluminum and antimony, may present some challenges. Achieving consistently high aluminum removal may require operation at two different pH levels because aluminum hydroxide is moderately soluble at a pH of 9.0 to 9.5. Antimony removal is generally accomplished through co-precipitation/adsorption with iron and will, therefore, be a function of the availability of iron for these reactions.

The footprint requirement for the HDS process is considerable. Clarifiers have to be designed with a low settling rate (less than 1 m/h) to obtain an effluent with low TSS.

Considering the characteristics of the water sources to treat, TSF water is the most suitable for HDS application. HDS is not considered the best option to treat the other three water sources or a combination of sources including a low percentage of TSF water. HDS could be considered for cases requiring treatment of a greater percentage of TSF reclaim water (see Section 4.4.4).

4.2.1.2 *Iron Co-precipitation / Adsorption*

Iron co-precipitation / adsorption is a chemical treatment process in which an iron salt is added to the water to be treated. Ferric sulphate, $Fe_2(SO_4)_3$, is normally used.

Iron co-precipitation is typically applied to target anionic forms of antimony, arsenic, molybdenum and selenium. The process is pH dependant with an optimal pH range of 4 – 6 for antimony [10] and molybdenum [2] removal. Operation at neutral pH (~7) is possible but some removal efficiency for antimony and molybdenum will be sacrificed.

Removal for specific parameters is also dependent on the form or oxidation state in which the element is present. Arsenic is better removed as arsenate (As V). Selenium is removed as selenite (Se IV). The process is not able to remove selenate (Se VI) (see Section 4.4.4.4). Oxidation / reduction may be required in combination with iron co-precipitation / adsorption to enhance removal for specific parameters.

Since this process can operate in the vicinity of the point of minimum solubility for aluminum hydroxide (minimum solubility occurs at a pH of ~6), good aluminum removal is expected. Although the process is not aimed at metals removal, some coincidental removal may occur by precipitation as oxides / hydroxides or through co-precipitation with iron.

Iron co-precipitation / adsorption is a proven technology which has been widely implemented at full scale throughout the industry. This process is able to reach a high removal for the target species: arsenic, molybdenum, antimony and selenium, dependant on speciation.

The process produces a relatively large quantity of sludge at low solids content. It may also occupy a large footprint, similar to HDS, depending on the specific equipment selected. The iron co-precipitation reaction requires a stirred reactor followed by a solid/liquid separation step. Solid/liquid separation can be performed using a conventional clarifier with a settling rate of roughly 1 m/h. However, this process is also amenable to other equipment choices which may reduce the space requirements. A lamella clarifier can reduce the total clarification surface area by using a stacked set of inclined plates to provide an effective settling area equivalent to a conventional clarifier in a more compact area. Lamella clarifiers are commonly designed with a rise rate of approximately 10 m/h. Lamella clarifiers are generally only applicable where the feed concentration of total suspended solids (TSS) is lower than 1000 mg/L and the feed does not have a tendency to produce scale; both of these conditions are satisfied for iron co-precipitation. High rate clarifiers (HRCs) are an extension of the lamella clarifier family. HRCs typically combine the reactors and settling steps within a single, compartmentalized vessel or a modular unit. They may also introduce an inert ballast material to increase the settling rate for precipitated or flocculated solids. For instance, ACTIFLO® advertises a rise rate of 40 m/h.

The main HRC competing technologies can be grouped as follows:

- Ballasted flocculation/sedimentation:
  - ACTIFLO® by Veolia Water Systems (silica sand ballasted flocculation with lamella clarification) and
  - CoMag™ by Evoqua (magnetite ballasted flocculation and clarification).
- High rate solids contact clarification/thickening:
  - DensaDeg® by Degremont (flocculation with chemically conditioned sludge recycle with settling tubes clarification).

HRC are a well proven technology with hundreds of installations around the world serving the municipal (drinking water and sewage) and industrial markets. It is also a proven technology in the mining industry (see Section 4.2.3).

Of these technologies, ACTIFLO® by Veolia and DensaDeg® by Degremont are most prominent in the mining effluent market. Evoqua is attempting to compete with the recent release of their CoMag™ system but there are relatively few installations in the mining sector, none at full-scale. One pilot-scale base metal demonstration and one pilot-scale precious metal demonstration have been publicized, both in the mid-western USA.

ACTIFLO® produces a dilute slurry waste containing approximately 0.5 to 1% dry solids. In contrast, DensaDeg® produces a more highly concentrated sludge (~5%). In either case, a thickener or other slurry dewatering technology may be required depending on the destination for the solid waste.

Since all the HRC systems are equipped with lamella plates, scaling issues when treating waters high in calcium, magnesium, carbonate (HCO_3) and sulphate (e.g., TSF reclaim water) may appear. DensaDeg® is a little less susceptible to scaling due to sludge recirculation. A modified ACTIFLO® version has been developed to overcome these issues. Neither HRC would be recommended for direct treatment of TSF reclaim water.

HRCs have a good flexibility to handle variable flows while obtaining an effluent with low TSS. Ability to manage flow variation is a weakness for conventional clarifiers. Flow to the Donlin Gold WTP is expected to vary since two of the sources (Wells and SRS) have limited local storage and the surface water sources (Upper CWD and TSF) will be influenced by climate and precipitation.

The feed water to the WTP will contain high concentrations of arsenic and antimony that will need to be reduced to allow the polishing step to reach the discharge objectives. All identified polishing technologies (Table 4-5) also require TSS removal as pre-treatment, which can be at least partly addressed using iron co-precipitation.

Suitability of this technology to address priority parameters (e.g., antimony and arsenic) in the feed water makes iron co-precipitation the recommended process for the initial stage of pre-treatment. Since the HRC equipment is more compact, well adapted to handle variable flows and the blend of waters to be treated has a low scaling potential, an HRC option is recommended in preference to a conventional clarifier with tank reactors.

4.2.1.3 *Ferrous Hydroxide Reduction*

Ferrous hydroxide reduction is a selenium removal process developed by the US Bureau of Reclamation. It is a two step process: reduction of selenate (Se VI) to selenite (Se IV) by Fe^{2+} followed by physical adsorption or iron co-precipitation of selenite (Se IV) by ferrihydrate or ferric hydroxide. Reduction and subsequent adsorption is pH dependant. The optimal pH range is 8 – 9. Selenium removal below 5 ppb has not been proven.

The technology has been demonstrated for selenium removal from effluent in other sectors (e.g., flue gas desulphurization).

For Donlin Gold, high operational costs for chemicals would be anticipated due to competition from oxygen, nitrate and sulphate in the reduction reaction.

This technology is not considered suitable for treatment to remove selenium from the Upper CWD water, which is the main source of selenium at Donlin Gold, due to high sulphate presence that will produce high chemical consumption.

4.2.1.4 *Electrocoagulation*

Electrocoagulation (EC) is a physical adsorption process where no reagent is required. Ferrous iron is produced by applying an electrical current to the water.

EC is potentially able to remove a wide range of contaminants (antimony, arsenic, molybdenum, selenium, heavy metals, TSS) by a combination of iron co-precipitation and redox reactions. The mechanism for selenium removal is similar to ferrous hydroxide reduction (see Section 4.2.1.3).

EC is more efficient when low concentrations of pollutants are present. Reduced sludge production is possible compared to other systems using iron co-precipitation.

EC should be considered to be an unproven technology. The technology has only been tested at pilot-scale outside of the mining sector and mainly with a focus on removal of biodegradable organics (i.e., BOD).

Due to the high risk involved in using a technology which has not been demonstrated, electrocoagulation has been discarded.

4.2.1.5 *Multimedia Filtration*

Media filtration achieves TSS removal primarily via size exclusion, but also via adhesion and adsorption. It is a proven technology used either as a polishing step after clarification or used directly for low turbidity effluents as a pre-treatment to ion exchange (IX) or reverse osmosis (RO).

Sand (quartz, silica) and anthracite are the most common filter media; however, garnet, magnetite and other media are also used. In multimedia filtration (MMF), multiple types of filter media are employed in layers in a single vessel with coarse, less dense media at the top of the filter and finer, more dense media at the bottom of the filter.

Normally, no reagent is required, but filter aid (e.g., flocculant polymer) may be added to agglomerate very fine particles. The typical filtration rate (or service flow) for MMF is 15 $m^3/m^2/h$ (15 m/h).

Periodically, the media filters must be taken out of service to undergo backwashing to remove the build-up of suspended solids. The backwash frequency depends on the TSS particle size and loading to the filter and filter design. In general, water recovery higher than 95% is obtained.

Depending on loading (i.e., feed TSS), solids characteristics and media properties, a filtrate TSS of less than 5 mg/L can readily be achieved via media filtration. MMF is typically effective for filtration of particles greater than 20 microns in diameter. Flocculation or additional fine filtration (e.g., 5-µm or 1-µm cartridges) may be required to obtain lower SDI values before RO treatment.

MMF is normally not recommended for applications where feed TSS may be greater than 125 mg/L or the sustained TSS concentration will be above ~75 mg/L.

Media filtration is a simple and robust technology which is frequently applied for effluent treatment across all mining sectors/subsectors.

Because manganese is in a reduced form in the feed water, oxidation and removal will be required. Therefore, greensand filtration (see Section 4.2.1.6) is preferred over simple particulate removal by MMF.

4.2.1.6 Manganese Greensand Filtration

Manganese greensand filtration is an adaptation of the multimedia filtration process that uses a special media covered with manganese oxide groups.

It is intended for dissolved iron and manganese removal to levels down to 0.05 mg/L. Some arsenic may also be removed along with any iron removal. An oxidant such as potassium permanganate ($KMnO_4$) or sodium hypochlorite (NaOCl) is required.

$KMnO_4$ is preferred due to a faster oxidation kinetics than NaOCl. Depending on the iron and manganese concentration in the feed, $KMnO_4$ is added either continuously or intermittently. Continuous addition is used for high iron and manganese concentrations. In this case, the oxidation of iron and manganese is achieved upstream of the filters and the filter acts as a conventional filter with the greensand available as a buffer to stabilize the process. With intermittent addition, the iron and manganese are adsorbed on the manganese media and the filter is regenerated with $KMnO_4$ upon exhaustion.

The maximum filtration rate for greensand filtration is between 10 – 15 m/h depending on the regeneration scheme applied. When continuous regeneration is used, a maximum filtration rate of 10 m/h is recommended. When concentrations of iron and manganese in the feed water are low, intermittent regeneration and a filtration rate of 15 m/h may be used.

Like any media filtration process, backwash is required. Like MMF, greensand media filtration is limited to a maximum sustained TSS concentration of ~75 mg/L.

Greensand filtration is a proven technology. It is widely used for pre-treatment of RO fed from well sources and in the direct treatment of groundwater streams for drinking water purposes.

Since high iron and manganese concentrations are expected in the feed water (see Section 4.1.2), greensand filtration is preferred to conventional media filtration, making this technology a good candidate to be used after initial solids removal by iron co-precipitation. Continuous addition of potassium permanganate will be used to cope with the relatively high iron and manganese concentration in the feed water.

4.2.1.7 *Microfiltration / Ultrafiltration*

Ultrafiltration (UF) and microfiltration (MF) are membrane-based technologies that achieve TSS removal via size exclusion. UF and MF are most commonly used for polishing after upstream TSS removal via other methods but can be used directly for low turbidity effluents.

They are particularly effective on fine and colloidal TSS. UF and MF technologies are quite similar, but have different pore size ranges and, therefore, have different particle size cut-off for TSS removal; while UF can remove TSS larger than 0.001 µm to 0.1 µm, MF can remove TSS above 0.1 µm to 10 µm size.

Membranes are available in many materials and in different configurations, including tubular, hollow fibre, spiral-wound (UF only), pleated cartridge (MF only), and flat sheet. The hollow fibre configuration is the most common for industrial effluent treatment. Low pressure (dead end configuration) or vacuum (submerged membranes) is used to drive effluent through the membranes to the permeate side while TSS greater than a specific size is retained on the reject side.

Chemical addition is not required in order for MF / UF to remove TSS. However, chemicals (flocculants and coagulants) can be used to enhance TSS removal, especially for fines and colloidal particles. Chemicals such as ferric salts and oxidants can also be used along with MF / UF filtration to promote arsenic and manganese removal. Submerged membranes with air scour are more suitable for applications using chemical reagents.

Periodically, the membranes must be taken out of service to undergo short backwashes (via flow reversal or with air) to remove the build-up of suspended solids and restore membrane flux.

The frequency of backwash depends largely on the feed water quality, the type of membrane and membrane configuration. Typically, a backwash occurs every 30 minutes and backwash duration is 30 seconds.

Less frequent chemically-enhanced backwash and/or chemical cleaning cycles ("clean-in-place" or "CIP") are also required to counter membrane scaling/fouling and restore flux. Replacement of membranes is required when membranes reach the end of their operational life.

MF / UF is an excellent pre-treatment for RO and IX and can produce water with a SDI < 1. The low feed SDI allows an RO system to be designed at a higher flux (20% higher), resulting in cost reduction. MF / UF can be used either as a polishing step after media filtration or as a stand-alone filtration system to replace other filtration steps. Permeate flux design and water recovery depends on feed water, with higher flux and water recovery possible for cleaner feed water. More than 90 - 95% water recovery is typical. UF membranes are being used more extensively as RO pre-treatment in drinking water and process water treatment. In the mining industry, and in particular with Barrick, UF is being incorporated ahead of RO (Lagunas del Norte and Pierina, Peru).

UF technology is considered a good candidate for clarification and / or filtration polishing ahead of the polishing step of the AWT plant.

4.2.1.8 *Ion Exchange (Selenium Removal)*

Ion exchange (IX) is the adsorption of ions of a given charge (either cationic or anionic) from a solution onto the surface of a solid material (ion exchange resin) and replacement in the solution by an equivalent ionic charge from the solid material surface. The resin is insoluble in water but hydrated and may be a salt, acid or base. Resins are formed through a polymerization process and functional groups added in various activation methods. A wide variety of functional groups can be grafted onto these polymers which are then used to selectively uptake ions. The functional groups of the polymers are initially associated with ions such as sodium (Na^+), hydrogen (H^+), chloride (Cl^-) and hydroxide (OH^-). The functional groups and resin polymer dictate what ions can be exchanged by the resin. Many resins are designed to target and selectively remove one or only a few parameters. To achieve the removal of a wide suite of parameters, several ion exchange steps in series, each with a specialized resin to target certain parameters may be required.

IX is a batch process. As the effluent passes through the column, functional group-associated ions are displaced from the resin by the targeted ions due to the higher affinity of the resin for the targeted ions. When resins become saturated with the targeted ions, they are regenerated by passing concentrated solution(s) ("regenerant") containing the replaced ions (e.g., NaOH, H_2SO_4, NaCl, HCl) through the column. Spent regenerant exits the column with the targeted ions in solution and may require further treatment prior to permanent disposal. The column is then flushed to remove residual regenerant and effluent treatment can re-commence.

Strong Base Anion (SBA) resins are able to remove anionic species (antimony, arsenic, molybdenum and selenium) from wastewater. However, selenium removal is influenced by oxidation state with Se (VI) being removed better than Se (IV). SBA resins are able to reach selenium concentrations as low as 0.01 mg/L on mine effluent [5]. Less performance data is available to relate the IX removal of other metal anions to the oxidation state although these may also be influenced in a similar way as selenium.

Selenium-containing effluent often also contains sulphate at much higher concentrations than selenium. This is the case for Upper CWD source water and any combination of waters containing Upper CWD water. SBA resins do not selectively remove selenate in preference to sulphate, so achievement of good selenium removal will also require a significant majority of the sulphate to be removed. Thus, the presence and concentration of sulphate in water to be treated has a major impact on the economics of ion exchange for selenium removal, as it impacts system sizing and regeneration frequency. Because of this, IX is not typically economically viable for selenium removal from mining effluent.

BioteQ has developed the Selen-IX™ process which uses SBA resin in sulphate form [4]. By operating the resin in sulphate form, the BioteQ process is able to remove selenium from solutions containing high background sulphate.

This technology can be considered for further evaluation as an option to pre-treat Upper CWD water alone. It may also be considered as a polishing step for the combined water sources (see Section 4.2.2.3).

4.2.2 Polishing Technology Options

4.2.2.1 Ion Exchange (TDS Removal)

Conventional water demineralization by IX uses a combination of steps to produce low TDS water. Typically, a cation step by using a strong acid cation (SAC) resin is followed by an anion step using SBA resin. An intermediate step to eliminate carbonate and reduce the load to the anion step is normally used.

IX demineralization is a proven technology widely used to produce process water. IX demineralization is cost effective if TDS is lower than 500 mg/L. For higher TDS, reverse osmosis (RO) is typically utilized instead of ion exchange.

Within this range (TDS lower than 500 mg/L), water recovery higher than 90% is possible. Chemical consumption for regeneration can be relatively high. Concentrate from regeneration may need treatment (will contain high sulphate or chloride depending on acid used for regeneration).

The WTP feed water (see Section 4.1.2) will have a 50th percentile TDS around 1000 mg/L and hence economics will not be favorable.

For the expected TDS concentration in the feed to the polishing treatment system, IX is not expected to be as cost-effective as Reverse Osmosis.

4.2.2.2 Ion Exchange (Metals Polishing)

Selective chelating IX resins are available and proven for water treatment and mineral processing applications targeting low discharge concentrations or high recovery for metal species such as cadmium, copper, mercury and nickel. However, selective resins are, by their nature, largely effective for only a small group of elements. Furthermore, in treatment of complex effluents with several competing ions present, removal efficiencies must be considered uncertain. To significantly reduce the concentrations of a number of different parameters in complex effluents, multiple ion exchange steps in series using a number of different resins that selectively remove parameters of concern may be required.

Ion Exchange may be applicable as a supplemental polishing step for a limited quantity of species where additional treatment is required downstream of principal polishing.

4.2.2.3 *Ion Exchange (Selenium Removal)*

Conventional IX process using SBA resins or specifically designed process (Selen-IX™) may be applied for selenium polishing and for more general polishing of anionic species like antimony, arsenic and molybdenum.

Applying IX treatment as a final polishing step rather than in pre-treatment may overcome all drawbacks previously discussed in Section 4.2.1.8 (high sulphate competition, high chemical consumption, low water recoveries).

Concentrated regeneration wastewater containing relatively high levels of selenium may need treatment if this waste cannot be used in the process or discharged directly to the TSF.

The ability to reach low selenium levels (less than 5 ppb) has been proven at pilot scale with the Selen-IX™ process [4]. Selenium removal using conventional IX technology has also been demonstrated at pilot scale for mining wastewater at Kennecott Mining Company [5].

The Selen-IX™ process includes a step to treat the brine while recovering the regenerant [4]. Although regenerant recovery adds considerable upfront capital cost and requires an expanded footprint, this may be an attractive option to consider if selenium treatment is required for Donlin Gold. Regenerant recovery permits significant reductions for the annual purchase cost for chemicals, including freight, and for hazardous chemical storage volume at site. Regenerant recovery also avoids potential issues related to accumulation of salts in the TSF.

Ion Exchange resin polishing is a relatively unproven technology for selenium removal to low discharge concentrations (<5 ppb) but the fundamentals are well understood and have been demonstrated in other applications. IX polishing offers one of the best approaches for reaching the selenium discharge objective. Hence, it will be considered for further evaluation although testing would be recommended prior to implementation.

4.2.2.4 *Reverse Osmosis*

Reverse Osmosis (RO) is a membrane filtration process which operates at high pressures allowing clean water (permeate) to flow across a membrane barrier while concentrating metals and salts in a reject brine stream (concentrate). The brine may require further treatment before disposal.

RO removes total dissolved solids (TDS) through the high pressure application of feed to selectively permeable membranes that exclude dissolved solids but allow the passage of water molecules. Removal performance is variable according to the species (refer to Table 4-14) but is typically well in excess of 90% with few exceptions (e.g., ammonia and WAD cyanide). Removal performance for ammonia and WAD cyanide may be influenced by choice of the operating pH. Removal efficiency is better for trivalent and divalent species (more than 98%) than for monovalent or uncharged species.

High feed pressure is required to overcome the osmotic pressure that develops across RO membranes, pushing the water molecules through the membrane to generate a low TDS permeate stream. Dissolved solids are retained on the feed side of the membrane and exit the pressure vessels as a brine stream having a high TDS concentration. RO treatment is only applicable to feed solutions where the osmotic pressure exerted by the fluid can be overcome at an operating pressure that will not damage the membranes (<6900 kPa [1000 psi]). Thus, RO treatment is limited to feed streams with TDS below approximately 45,000 mg/L.

When brackish water is treated (TDS less than 5000 mg/L), water recoveries between 50% and 75% are possible, depending on the scaling potential of the feed water. Antiscalant and pH adjustment are normally used to reduce the potential for scale deposition and membrane fouling.

A clean-in-place (CIP) system is required to restore membrane flux when loss of permeate flux, rejection or high differential pressure occurs due to fouling or scaling. CIP frequency depends on pre-treatment and membrane flux design and the RO operating conditions. A low SDI is required in the RO feed water to prevent membrane fouling. SDI values lower than 3 are recommended.

RO is a well proven technology used widely to produce drinking water and high quality process water. Use in the mining industry is growing, driven by enforcement of increasingly stringent discharge criteria. RO is widely viewed as a blanket treatment solution and its ability to reject multiple contaminants to very low levels, if adequate pre-treatment is provided, makes this technology attractive for Donlin Gold.

The high volume of brine which is produced by an RO system is normally a drawback for this technology since outlets for this concentrated wastewater are not always available. However, for Donlin Gold, the projected water quality of the brine allows it to be reused in the process plant [12]. Soluble salts are not expected to accumulate in the TSF to concentrations which would be detrimental for the process.

RO is retained as an option for the polishing step to be used in the AWT plant.

4.2.2.5 Nanofiltration

Nanofiltration (NF) is a membrane process similar to reverse osmosis that removes multivalent dissolved solids (including aluminum, arsenic, copper, iron, lead, manganese, nickel, radium-226, selenium and zinc) from a feed via the pressurized application of feed water to a charged, selectively permeable membrane that rejects multivalent ions. Unlike reverse osmosis, however, the membrane allows the passage of water and monovalent ions (e.g., ammonia, chlorides, cyanide, sodium, etc.). The required feed pressure for NF is lower than for RO.

Nanofiltration membranes exclude multivalent dissolved solids by charge and size. Like RO, NF generates a low TDS permeate stream and a concentrated, high TDS brine stream on the feed side of the membrane.

Nanofiltration has been proposed as a substitute to reverse osmosis in instances where rejection of multivalent ions (e.g., Cu^{2+}, Ni^{2+}) is more important than rejection of monovalent ions and where lower removal efficiencies are acceptable.

Since there is a need for high removal of monovalent species (i.e., ammonia, WAD cyanide) in the Donlin Gold application, NF is considered unlikely to reach the discharge objectives and has been removed from further consideration.

4.2.2.6 Evaporation and Crystallization

Evaporation is typically utilized when the feed TDS concentration is above 35,000 mg/L or where the feed temperature is high and/or a waste heat source is available. Evaporators can produce a treated water (distillate) that is superior to RO permeate, containing less than 10 mg/L TDS. High water recovery (>90-95%) can also be obtained, overcoming some of the RO disadvantage with respect to brine management. Evaporation is frequently combined with crystallization to evaporate the remaining water in the evaporator brine, producing a dry salt crystal for disposal and providing a zero liquid discharge treatment solution. The capital and operating (energy) costs for evaporation are high. RO pre-treatment is often applied for pre-concentration of the feed to minimize the required capacity of the evaporator for installation.

Evaporation is a proven and robust technology, with less reliance on pre-treatment in order to achieve stable operation.

Since dilute streams containing only 1000 – 1200 mg/L TDS are projected for Donlin Gold, the energy required to produce a concentrated brine would be very high. Thus, this option is not feasible as a stand-alone treatment solution. Although evaporation will not be considered as a primary treatment option, it could be considered if RO brine requires treatment to prevent salt build-up in the TSF. In this way, the size of the crystallization unit would be reduced.

4.2.3 Technology and Equipment Selection

4.2.3.1 Pre-Treatment

Iron co-precipitation has been recommended as the initial treatment step to reduce aluminum, antimony, arsenic and TSS concentrations. Iron co-precipitation can be implemented using conventional tank reactors and a clarifier or with high rate clarification (HRC) equipment. Compared to conventional clarifiers, HRC provides some advantages in this application which ultimately lead to lower project cost. Table 4-6 presents a comparison between HRC and conventional clarifiers. On the basis of the comparison in Table 4-6, HRC equipment is proposed for Donlin Gold.

Table 4-6: Comparison Between HRC and Conventional Clarification

Property	HRC	Conventional Clarification
Extent of Pre-Assembly	Pre-assembled and integrated coagulation, flocculation and clarification	Separate tank reactors and clarifier
Installation Cost	Lower	Higher (site-erected vessels)
Design Rise Rate (m/h)	30 - 40	1
Footprint (m^2)	70	1000
Feed TSS – Max. (mg/L)	1000	Essentially no limit
Underflow % Solids (wt%)	0.2 - 5	2 - 3

Additional pre-treatment for removal of manganese and TSS is needed following HRC iron co-precipitation and in advance of the polishing treatment. Greensand media filtration (with downstream 5-µm absolute cartridge filtration) is recommended for this purpose in preference to ultrafiltration (UF). This pre-treatment is deemed sufficient to ensure reasonably trouble-free operation and low membrane cleaning and replacement frequency for the downstream RO. UF coupled with chemical conditioning to precipitate manganese is a viable alternative to the greensand filter. It would generally be acknowledged that UF will produce a permeate with lower SDI than the filtrate from the greensand filters and SDI normally correlates well with RO membrane fouling. However, both media filters and UF are accepted forms of pre-treatment for RO. Table 4-7 presents relative pros and cons for each technology.

Table 4-7: Comparison Between Greensand Filtration and Ultrafiltration

Property	Greensand Filtration	Ultrafiltration
Water Recovery	Higher (>95%)	Lower (~90%)
Footprint	Marginally higher	Marginally lower
SDI	Higher (<5)	Lower (<1)
RO Design	Standard	High flux (minor cost saving)

4.2.3.2 *Polishing*

Reverse osmosis has been recommended to perform the polishing treatment in the Donlin Gold AWT plant. None of the other polishing technologies reviewed in Section 4.2.2 appear to be cost competitive with RO while achieving compliance with the discharge objectives. RO is suitable for Donlin Gold for several reasons:

- RO is able to achieve good removal for all parameters requiring treatment.
- TDS for the blended feed is relatively low allowing high RO recovery without concern for membrane scaling or fouling; and
- RO brine is suitable for use in the process and will not cause significant deterioration of the TSF reclaim water quality [12].

Final polishing using IX for specific contaminants has not been retained for the base AWT design since it is deemed to be unnecessary at the start of operations and may not be required at all, depending on several other factors. Parameters which represent the greatest risk to compliance are discussed in Section 4.4.4.

4.2.3.3 Precedents for Recommended Technologies and Equipment

Table 4-8 provides a listing of some precedents for the use of the recommended technologies and equipment.

Table 4-8: Selected Technologies – Precedents

Technology	Pre-Treatment	Polishing	Precedents
Iron Co-Precipitation with HRC	X		Barrick Williams Mine, Ontario, Canada: • ACTIFLO® – 2,000 m^3/d – Treatment of tailings water for TSS, molybdenum and antimony removal. Bendigo Mining New Moon Gold Mine, Australia: • ACTIFLO®, Greensand Filtration and Reverse Osmosis – 7,000 m^3/d – Treatment of underground mine water [20]. Goldcorp Eleonore Mine, Quebec, Canada: • ACTIFLO® – Treatment of tailings water for metals and arsenic. Goldcorp Red Lake Mine, Ontario, Canada: • ACTIFLO® – 30,000 m^3/d – Arsenic abatement from tailings pond discharge. Vale Long Harbour Nickel Processing Plant, Newfoundland, Canada: • ACTIFLO® – 25,000 m^3/d – Treatment of tailings water. Argo Tunnel AMD Treatment System, Colorado, USA: • DensaDeg® and filtration polishing – Flow unavailable – Active and semi-passive lime treatment of AMD [8]. Diavik Diamond Mine, Northwest Territories, Canada: • DensaDeg® – 15,000 to 90,000 m^3/d – Treatment of underground mine water.
Greensand Filtration	X		Common technology in drinking water treatment for iron and manganese (e.g., approved technology in Quebec, Canada [14]).

Technology	Pre-Treatment	Polishing	Precedents
Reverse Osmosis		X	• Barrick Lagunas del Norte – Peru – AMD treatment. • Barrick Pascua Lama – Chile – AMD treatment. • Barrick Pierina – Peru – AMD treatment. • Barrick Richmond Hill Mine – South Dakota, USA – Selenium polishing from mine water after pre-treatment by iron reduction and precipitation [19]. • Newmont – Peru – Gold leaching operation; flow: 340 m^3/h [15]. • Rio Tinto Kennecott South Site – Utah, USA – Bingham Canyon Water Treatment Plant – Two RO trains; flow: 760 m^3/h each [19].

4.3 Process Description

The process description has been based on preliminary design work. Equipment quantities, capacities and reagent use may change during detailed design.

The wastewater coming from the four individual sources will be collected and blended in the WTP feed tank. This tank will have a volume of 265 m^3, which will allow for a hydraulic retention time of 15 minutes at peak flow. Air for mixing will be supplied by means of two blowers (one duty, one stand-by) rated for 500 Nm^3/h each. In addition to mixing, the air will allow for iron oxidation.

From the feed water mixing tank, three pumps (two duty, one stand-by) will feed the high rate clarifiers. The HRC feed pumps will be 530 m^3/h each.

Sulphuric acid and ferric sulphate will be dosed in line ahead of the HRC to adjust pH for the iron co-precipitation process. The pH setpoint and ferric sulphate dosage will be adjusted to optimize antimony removal. The operating pH is expected to be in the range of 5 – 6 with a ferric sulphate dosage of approximately 120 mg/L.

After chemical dosing, the influent enters the high rate clarification step. There will be two HRC. Each HRC will be designed to treat 530 m^3/h. The two HRC will operate in parallel and there will be no redundancy. A single HRC will be capable of operation at 150% of the design flow during periods where the other HRC unit is out of service.

In the HRC, a polymeric flocculant will be added to assist with the agglomeration of the precipitated ferric hydroxide and co-precipitates. The solids are separated in the clarification step. The sludge produced in this step is sent to the Backwash Wastewater / Clarifier Sludge Receiver Tank, which will also receive the wastewater produced during the backwash of the greensand filters. Sludge flow from the clarifiers is estimated at 3% of the peak flow and will not vary with HRC throughput.

The removal efficiency for antimony in the HRC at the proposed operating conditions is expected to be 75 – 80% [10]. Greater than 95% arsenic removal is also anticipated in the HRC at these conditions. Aluminum and molybdenum removal will also occur.

The overflow from the HRC clarifier will be collected in the Clarified Water Transfer Tank. This tank will have a capacity of 88 m^3. The hydraulic retention time will be 5 minutes at peak flow.

From the Clarified Water Transfer Tank, water will be pumped to the greensand media filters by means of four pumps (three duty, one stand-by). The pumping capacity will be 343 m^3/h per pump. There will be three parallel trains of greensand media filters which will each be fed by a dedicated pump. Each train will be made up of four 3.66 m (12-foot) diameter filters.

No redundancy has been allowed for filtration. With all filters in operation, the filtration rate will be 8 m/h. The maximum filtration rate for the on-line filters in a single train will be 10.7 m/h with one of the filters off-line for backwashing. This rate is slightly higher than the recommended rate of 10 m/h for greensand media filters with continuous addition of $KMnO_4$ (see Section 4.2.1.6) but is judged to be acceptable.

The greensand media filters will be dual media filters. The top layer will be anthracite intended for TSS removal and the bottom layer will be the greensand media itself. Potassium permanganate ($KMnO_4$) will be injected upstream of the greensand filters. The $KMnO_4$ will oxidize the manganese, which will be in the +2 oxidation form. Consumption of $KMnO_4$ for oxidation of iron is not expected because the iron should already be fully oxidized by aeration in the WTP feed tank. The expected dosage of $KMnO_4$ will be between 3 – 5 mg/L.

The pH at the inlet of the greensand filters may be as low as 5 depending on the operating pH needed in the HRC to achieve target antimony removal. At this low pH, oxidation kinetics will be affected. Sodium hydroxide (NaOH) will be added in-line to raise the pH to 7 before the injection of $KMnO_4$.

The greensand filters will be backwashed with air and water. The backwash system will comprise two blowers (one duty, one stand-by) with flows of 315 Nm^3/h each and two backwash pumps (one duty, one std-by) with flows of 378 m^3/h each.

Assuming an average feed of 5 mg/L of TSS and 1 mg/L of manganese, a daily backwash frequency is expected at maximum flow. Typical backwash times are 2 – 3 minutes for air backwash followed by 10 minutes water backwash. Concentrate from the RO will be used for backwash.

Wastewater from filter backwash will be sent to the Backwash Wastewater / Clarifier Sludge Receiver Tank. This tank will have a capacity of 95 m^3 and will be able to store one filter backwash while receiving the continuous sludge flow coming from the HRC. The tank will be equipped with a 5 HP mixer intended to keep solids in suspension.

The combined wastewater from the HRCs and the greensand media filters will be sent to the TSF by means of two pumps (one duty, one stand-by) with a design flow per pump of 100 m^3/h.

The filtrate from the greensand filters will feed the RO step. RO pre-treatment will be required to protect the membranes from oxidation, fouling and scaling.

Since the RO membranes are susceptible to degradation by any residual oxidant, sodium metabisulfite ($Na_2S_2O_5$) will be used to remove any trace of $KMnO_4$. The expected $Na_2S_2O_5$ dosage should be in the range of 2 – 3 mg/L.

Antiscalant addition is also required to maintain sparingly soluble salts ($CaSO_4$, $CaCO_3$, $BaSO_4$, $SrSO_4$, SiO_2) in solution and prevent precipitation on the membranes. An antiscalant dosage of 5 mg/L is expected.

In order to protect the RO membranes from fouling, a 5-micron absolute cartridge filtration system will be installed ahead of the RO system. Polypropylene disposable cartridge filters will be used. The recommended cartridge replacement frequency is once every two to three months or whenever the pressure drop across the cartridge filter housing exceeds 103 kPa (15 psi).

The RO system will be made up of four parallel trains (three duty, one stand-by). Each train will be fed by two high pressure pumps (one duty, one stand-by). Each pump will be rated for 343 m^3/h at 15 bar.

The RO will operate at 75% recovery. The RO permeate will be collected in a RO permeate tank. Before entering the tank, the RO permeate will receive a dose of NaOH to adjust pH within the target range (7.5 – 8) for discharge. Soda ash (Na_2CO_3) may be added as an alternative or in addition to NaOH for pH adjustment and to also increase the alkalinity of the treated water.

The RO permeate tank will have a capacity of 125 m^3. The hydraulic retention time will be 10 minutes at peak flow. The tank is intended only as a break tank before transfer of the treated water to the discharge point. There will be two treated water transfer pumps (one duty, one stand-by). The flow for the treated water transfer pumps will be 800 m^3/h each.

In normal operation, treated water will be pumped to the discharge point. If, for any reason, the treated water is out of specification, then the water will be transferred to the Lower CWD until the problem is solved.

The brine from the RO process will be collected in a tank (RO Brine Water Tank). The capacity of this tank will be 95 m^3. The tank capacity includes 63 m^3 water storage for one greensand media filter backwash. Two sets of pumps will be installed to draw from this tank: one set for greensand media filter backwash and the other set (one duty, one stand-by) to transfer RO brine to the reclaim water tank for reuse in the process. The design flow for the RO brine transfer pumps will be 300 m^3/h each.

In normal operation, the RO membranes can become fouled by mineral scale, biological matter, colloidal particles and insoluble organic constituents. Deposits build up on the membrane surfaces during operation until they cause loss in normalized permeate flow, loss of normalized salt rejection, or both.

Elements should be cleaned when one or more of the below mentioned trigger conditions is satisfied:

- The normalized permeate flow drops 10%;
- The normalized salt passage increases 5 - 10%; or
- The normalized pressure drop (feed pressure minus concentrate pressure) increases 10 - 15%.

Expected cleaning frequency in a well designed and operated RO system should be quarterly.

A clean-in-place (CIP) system will be used. A typical CIP system consists of a chemical mixing tank (PP or FRP), an electric heater, a low pressure pump, a 5-μm cartridge filtration system and valves and piping to recirculate permeate and concentrate to the chemical tank.

Typical cleaning chemicals to be used will be hydrochloric acid (0.1% HCl solution strength) and sodium hydroxide (0.05% NaOH solution strength). RO permeate will be used for dilution of the cleaning chemicals and flushing of the RO system.

The waste from RO CIP will be sent to the TSF. No neutralization step has been considered to neutralize the acidic and alkaline effluent generated during RO cleaning. Neutralization and/or dilution is considered to happen in the TSF.

Table 4-9 summarizes the main equipment of the WTP. A flow schematic of the WTP is presented in Figure 4-1.

As shown in Appendix B, the same treatment process is essentially applicable for the treatment of any of the water sources individually. Partial bypass of the RO system may be possible if treating water predominantly from the Wells due to lower feed concentrations. Additional sodium hypochlorite dosing may be required if treating water only from the SRS or TSF to help manage the higher WAD cyanide and ammonia (TSF only) concentrations. Polishing treatment using IX may need to be added after an initial period of operation to respond to the anticipated increasing concentrations of selenium (Upper CWD) and mercury (TSF) over mine life if treating these water sources individually.

It is only for individual treatment of TSF water that the treatment process would incorporate a new element. For treatment of TSF water, it would be recommended to include a high density sludge (HDS) circuit as an initial treatment step to reduce the concentrations of sulphate, magnesium and most other metals before applying the usual treatment process. Individual treatment of TSF water may also require improved molybdenum removal by lowering pH and adding ferric sulfate before the MMF to reduce the molybdenum concentration via co-precipitation / adsorption mechanisms.

Table 4-9: WTP Main Equipment List

Equipment	Quantity	Characteristics
WTP feed tank	1	Material: Concrete HRT: 15 min (at peak flow) V: 265 m^3
WTP feed tank blowers	2 (one duty, one stand-by)	Flow: 500 Nm^3/h Head: 0.5 bar Power: 10 HP
HRC feed pumps	3 (two duty, one stand-by)	Flow: 530 m^3/h Head: 2 bar Power: 75 HP
HRC	2 (two duty)	Rise rate: 40 m/h (at peak flow) HRT: 25 min (at peak flow)
Clarified water transfer tank	1	Material: Concrete HRT: 5 min (at peak flow) V: 88 m^3
Greensand filter feed pumps	4 (three duty, one stand-by)	Flow: 343 m^3/h Head: 5 bar Power: 100 HP
Greensand filters	12 (four per train)	Diameter: 3.66 m Side straight height: 1.8 m Media: Anthracite/greensand
Greensand filter backwash pumps	2 (one duty, one stand-by)	Flow: 378 m^3/h Head: 2 bar Power: 40 HP
Greensand filter backwash blowers	2 (one duty, one stand-by)	Flow: 315 Nm^3/h Head: 0.5 bar Power: 7.5 HP
RO high pressure pumps	8 (one duty, one stand-by per RO train)	Flow: 343 m^3/h Head: 15 bar Power: 300 HP
RO	4 trains (three duty, one stand-by)	Flow: 343 m^3/h (per train) Recovery: 75% Permeate flux: 14 $USG/ft^2/day$ (at peak flow)
RO permeate water tank	1	Material: Concrete HRT: 10 min (at peak flow) V: 125 m^3
Treated water discharge pumps	2 (one duty, one stand-by)	Flow: 800 m^3/h Head: 3 bar Power: 125 HP

Equipment	Quantity	Characteristics
RO CIP system	1	Chemical mixing tank Material: PP or FRP V: 10 m^3 Heater: 60 kW Pump: 40 HP 5-µm cartridge filter
RO brine water tank	1	Material: Concrete V: 95 m^3
RO brine water tank transfer pumps	2 (one duty, one stand-by)	Flow: 300 m^3/h Head: 3 bar Power: 50 HP
Backwash Wastewater / Clarifier Sludge Receiver Tank	1	Material: Concrete V: 95 m^3
Wastewater transfer pumps	2 (one duty, one stand-by)	Flow: 100 m^3/h Head: 3 bar Power: 15 HP
Receiver tank mixer	1	Material: SS Power: 5 HP

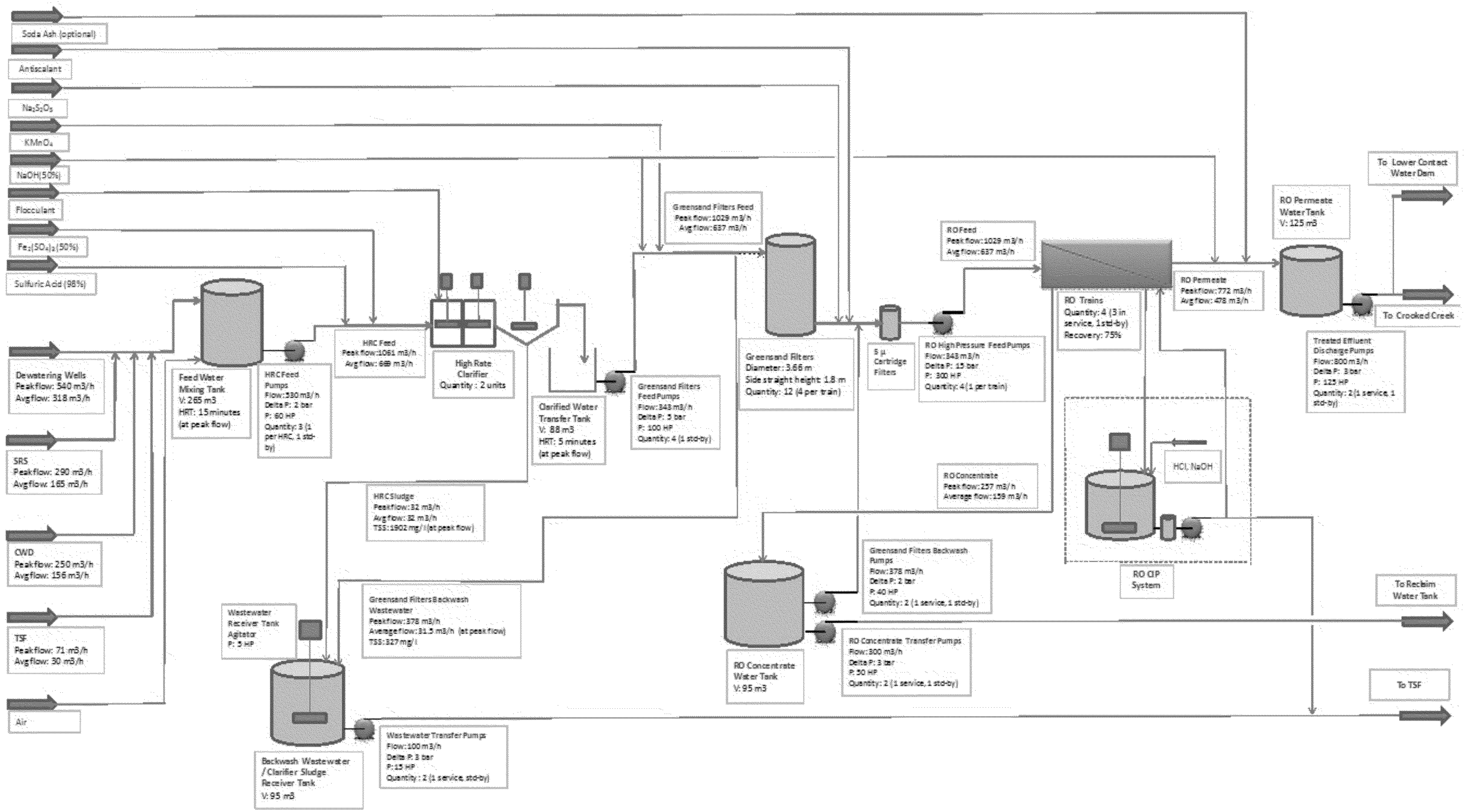

Figure 4-1: **WTP Flow Schematic**

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4.3.1 *Preliminary Lay-Out*

Figure 4-2 presents the WTP preliminary lay-out. The estimated plant footprint is 2200 m^2. If IX polishing becomes necessary to manage increasing feed concentrations for species such as selenium, it would be recommended to add this additional treatment equipment in an adjoining building. An IX polishing system is expected to require 800 to 1500 m^2. Considering that need for this treatment may never arise and the space required represents a substantial increase to the base case lay-out, it is not advisable to allow this space for potential future requirements within the initial building. However, selection of the WTP location must consider this potential need for future expansion.

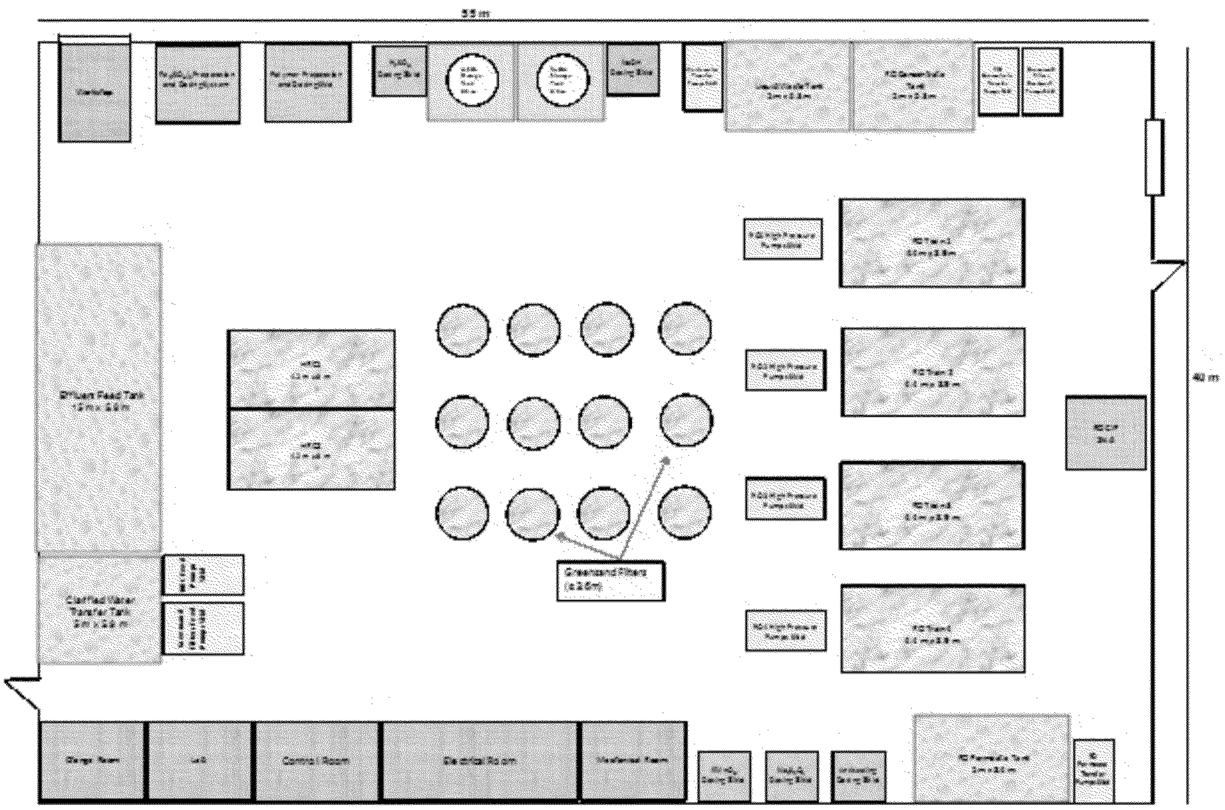


Figure 4-2: Preliminary Lay-Out

4.4 Advanced Water Treatment System Performance

4.4.1 Treated Water Quality

Table 4-10 presents the projected treated water quality with comparison to the discharge objectives. The assumptions for removal efficiency at each AWT unit operation, on which this projection has been based, are documented in Section 4.4.3. Only parameters which exceed the discharge objectives in the feed water and select other parameters of relevance for design (e.g., chloride) are presented. Treatment slides are provided in Appendix B showing the projected performance for the treatment system when processing water from the different sources individually. While the plant has been optimized for a combination of sources, as demonstrated, it is capable of treating dewatering wells and SRS water streams individually.

It must be emphasized that the projected treated water quality has not been based on bench or pilot testing and will require verification in future work.

Table 4-10: Treated Water Quality

Parameter	Treated Water (mg/L)	Discharge Objective (mg/L)
Aluminum	< 0.006	0.087
Ammonia	0.6	2.16
Antimony	< 0.004	0.006
Arsenic	< 0.004	0.01
Barium	< 0.02	2
Boron	< 0.04	0.75
Cadmium	3.6 E-05	2.7 E-04
Chloride	< 0.12	230
Chromium, total	< 0.0005	0.1
Cobalt	< 0.0003	0.05
Iron	< 0.005	1
Lead	0.0004	0.0032
Manganese	< 0.002	0.05
Mercury[1]	1.1 E-05	1.2 E-05
Molybdenum	< 0.005	0.01
pH	7 – 8	6.5 – 8.5
Selenium	< 0.004	0.005
Sulphate	< 22	250
TDS	< 38	500
WAD Cyanide[2]	< 0.0045	0.0052
Zinc	< 0.003	0.119

Notes:
[1] Mercury removal performance is based on 30 m^3/h TSF flow, 5 ppb mercury in the TSF, 1.3% TSF seepage contribution to SRS flow and 96% RO removal.
[2] WAD cyanide removal performance is based on 30 m^3/h TSF flow, 0.26 mg/L cyanide in the TSF and 80% RO removal.

4.4.2 *Waste Streams*

Table 4-11 presents a list of the WTP waste streams along with their flows, main characteristics and destination. The projected TDS and sulphate concentrations in the RO brine are below the steady-state concentrations in the TSF; this WTP waste stream can be accepted in the Reclaim Water Tank for direct use in the process plant [12]. Indicated values are for average flow to the WTP.

Table 4-11: Waste Streams

Waste Stream	Type	Flow or Volume	Main Characteristics	Destination
HRC sludge	Continuous	32 m^3/h	• TSS: 1902 mg/L • pH: 5 – 6 • Other parameters: similar quality to feed water	TSF
Greensand media filter backwash wastewater	Intermittent	504 m^3/day	• TSS: 327 mg/L • pH: 7 -8 • Other parameters: similar quality to RO concentrate	TSF
RO brine	Continuous	138 m^3/h	• TDS < 5070 mg/L • SO_4 < 2880 mg/L • Total Metals: < 2 mg/L • Se: <0.15 mg/L • Cl: < 15.6 mg/L • Ca: < 540 mg/L • Mg: < 490 mg/L • $NH_{4:}$ < 9 mg/L • CN_{WAD}: < 1 mg/L • pH: ~ 8	Reclaim Water Tank
RO CIP waste	Intermittent	100 m^3/cleaning cycle/RO train	• pH: 1 -12	TSF

4.4.3 *Advanced Water Treatment Performance Assumptions*

The projected treated water quality in Table 4-10 has been derived using a set of assumptions on the removal efficiency for various species across the different AWT unit operations. Table 4-12, Table 4-13 and Table 4-14 present the assumptions used for each stage of treatment. Removal efficiencies are only tabulated for parameters requiring treatment in order to meet the discharge objectives. References attributed to Hatch are based on data from confidential projects.

Table 4-12 presents the basis for determining the HRC outlet quality.

Table 4-12: Performance Assumptions: HRC System

Parameter	Inlet Conc. (mg/L)[1]	Outlet Conc. (mg/L)	Removal Efficiency (%)	References
Aluminum[2]	1.31	< 0.2	> 85	AWWA [1]
Antimony	0.149	< 0.04	75	Guo et al [10]
Arsenic	2.23	< 0.2	> 90	Hatch
Iron	5.4	0.5 - 1	82 – 91	Hatch
Manganese	1.1	0.55	50	Hatch
Molybdenum	0.052	0.050	< 5	Hatch / Aube [2]

Notes: [1]: Refer to 95th percentile blended feed (Table 4-3).
[2]: Aluminum outlet concentration based on solubility of aluminum hydroxide at pH 5.5 = 200 ppb. Solubility is 27 ppb at pH 6.0.

Table 4-13 provides the assumptions used for removal by the greensand media filters.

Table 4-13: Performance Assumptions: Greensand Media Filtration System

Parameter	Inlet Conc. (mg/L)	Outlet Conc. (mg/L)	Removal Efficiency (%)	References
Arsenic[1]	< 0.2	< 0.05	75	Hatch
Iron	0.5 - 1	0.05 - 0.1	80 - 95	MDDEP [14]
Manganese	0.55	< 0.05	> 90.9	MDDEP [14]

Notes: [1]: Combined removal efficiency of 98% assumed for arsenic (HRC + greensand).

The performance assumptions for the RO system are given in Table 4-14.

Table 4-14: Performance Assumptions: Reverse Osmosis System

Parameter	Inlet Conc. (mg/L)	Outlet Conc. (mg/L)	Removal Efficiency (%)	References
Aluminum	< 0.2	< 0.006	97	Hatch
Ammonia	2.74	0.6	80	Hatch
Antimony	< 0.04	< 0.004	90	Hatch
Arsenic	< 0.05	< 0.004	91	Hatch
Cadmium	5.9 E-04	3.6 E-05	94	Hatch
Lead	0.013	0.0004	97	Hatch
Manganese	< 0.05	< 0.002	97	Hatch
Mercury[1]	2.9 E-04	1.1 E-05	94 - 98	Hatch / PWG [16]
Molybdenum	0.05	< 0.005	91	Hatch
Selenium	0.038	< 0.004	91	Hatch / PWG [16]
Sulphate	720	< 22	97	Hatch
TDS	1267	< 38	97	Hatch
WAD Cyanide[2]	0.13 to 0.26	< 0.0045	80 - 90	Hatch / PWG [16]

Notes: [1] Mercury removal performance is based on 30 m^3/h TSF flow, 5 ppb mercury in the TSF, 1.3% TSF seepage contribution to SRS flow and 96% RO removal.
[2] WAD cyanide removal performance is based on 30 m^3/h TSF flow, 0.26 mg/L cyanide in the TSF and 80% RO removal.

4.4.4 Limiting Factors for Treatment

From Table 4-3 it is evident that there are multiple parameters in the blended worst-case (95^{th} percentile) WTP feed which exceed the discharge objectives. These parameters are:

- Aluminum (Al);
- Ammonia (NH_4);
- **Antimony (Sb)**;
- Arsenic (As);
- Cadmium (Cd);
- Iron (Fe);
- Lead (Pb);
- Manganese (Mn);
- **Mercury (Hg)**;
- Molybdenum (Mo);
- **Selenium (Se)**;
- Sulphate (SO_4);
- Total Dissolved Solids (TDS); and
- **WAD cyanide (CN_{wad})**.

The parameters identified in **bold font** in the above list represent the greatest challenges for treatment and will be discussed in detail in the following sections.

The projected treated water quality is shown in Table 4-10. The proposed design of the WTP is expected to be sufficient to produce treated water which complies with the discharge objectives.

4.4.4.1 Antimony

From thermodynamic equilibrium considerations, it is believed that antimony predominately exists as Sb(V) in oxic systems and mainly as Sb(III) under anoxic conditions. Antimony should be expected to be present mainly as Sb(V) at Donlin Gold.

In the pH range of 2–11, Sb(III) should form a neutral complex, $Sb(OH)_3$, whereas Sb(V) should exist as a negatively charged complex, $[Sb(OH)_6]^-$.

Antimony is removed by iron co-precipitation in a similar fashion as arsenic. Removal of Sb(III) is relatively insensitive to pH. However, removal of Sb(V) requires higher iron dosages and a lower operating pH by comparison to arsenic treatment. At a pH in the range 5–6, removal rates higher than 90% are possible.

At a pH of 6, a minimum iron dosage of 6 X 10^{-4} mol/L is required to reach 90% removal for Sb(V). Removal efficiencies as high as 95% are possible with the iron dosage increased to 8 X 10^{-4} mol/L. Data relating antimony removal efficiency with operating pH and iron dosage from laboratory scale experimentation is available in Guo et al [10].

It will be possible to achieve the discharge objectives for antimony, even in the case of individual treatment of Upper CWD water which has the highest antimony concentration of the four sources, by adjusting the iron dosage and operating pH for the HRC. These operating conditions will need to be monitored and trended against antimony removal to optimize this relationship.

4.4.4.2 *Cyanide*

Cyanide can be broadly classified into three categories: free cyanide, weak acid dissociable (WAD) cyanide and total cyanide. Free cyanide (HCN/CN^-) is the active form of cyanide used in the process for gold leaching. WAD cyanide includes free cyanide and weak and moderately strong cyanide complexes (e.g., $Ag(CN)^{2-}$, $Cu(CN)^{2-}$, $Cu(CN)_3^{2-}$, $Ni(CN)_4^{2-}$) which will decompose in mildly acidic solution (pH 3 to 6). Total cyanide includes WAD cyanide and more stable, strongly-bound cyanide complexes such as $Au(CN)^{2-}$, $Co(CN)_6^{4-}$ and $Fe(CN)_6^{4-}$. Thus, the concentration of WAD cyanide is always equal or less than the concentration of total cyanide.

The WAD cyanide concentration in the TSF reclaim water is a factor which potentially constrains the volume of TSF reclaim water which can be treated while ensuring compliance with the discharge objective of 0.0052 mg CN_{WAD}/L.

Process modeling (using Metsim®) and targets for operational control are based on the total cyanide concentration, not WAD cyanide. The WAD cyanide concentration in the TSF reclaim water is currently estimated based on the following approach:

- The tailings slurry after CN destruct is designed to target a total cyanide concentration of less than 10 mg CN_{TOT}/L at a flow of 810 m^3/h (flows from Metsim®).
- Blended flow from the plant to the TSF is composed of:
  - Tailings slurry after CN destruct: 810 m^3/h @ 10 mg CN_{TOT}/L.
  - Overflow from chloride wash CCD: 1630 m^3/h @ 1.3 mg CN_{TOT}/L
  - Underflow from flotation tails thickener: 1840 m^3/h @ 1.3 mg CN_{TOT}/L.
- Total cyanide concentration in the blended flow from the plant to the TSF is 3.0 mg CN_{TOT}/L.
- The TSF total cyanide concentration is diluted slightly by inflow from natural sources. Based on BGC's water balance, inflow from annual precipitation and runoff is approximately 280 m^3/h versus flow of blended tails from the plant at 3971 m^3/h. This results in a total cyanide concentration of 2.6 mg CN_{TOT}/L.

- Natural degradation is assumed to reduce the total cyanide concentration in the TSF reclaim to 50% of the original value. The resulting total cyanide concentration is 1.3 mg CN_{TOT}/L.
- The concentration of WAD cyanide is conservatively assumed to be equal to the concentration of total cyanide. Therefore, the WAD cyanide concentration in TSF reclaim water is assumed to be 1.3 mg CN_{WAD}/L. This assumption is perhaps overly conservative given that the CN destruct circuit does not destroy ferrocyanide (FeCN) complexes, which would thus be expected to build up in the TSF, causing total cyanide to be greater than WAD cyanide.

The water treatment process currently assumes that no cyanide is removed by the high rate clarifier or greensand filter. All cyanide removal is accomplished by the RO. Cyanide rejection of 80% for the RO has been assumed based on the conservative assumption that the WAD cyanide is entirely free cyanide. Rejection is expected to be higher for cyanide in other forms.

An average flow from other sources to the WTP of 639 m^3/h has been estimated (refer to Table 4-2). The 95^{th} percentile WAD cyanide concentration in these blended streams is currently calculated to be 0.012 mg CN_{WAD}/L, which is only slightly higher than the discharge limit, so it is only through introduction of TSF reclaim water that cyanide becomes a constraint for the WTP.

Based on 80% cyanide removal in the WTP, the maximum WTP feed concentration for WAD cyanide which can be accepted is 0.026 mg CN_{WAD}/L. At a cyanide concentration of 1.3 mg CN_{WAD}/L in the TSF reclaim water, it is possible to treat just 7 m^3/h of TSF water while meeting the 0.0052 mg/L discharge objective. However, as discussed above, the assumption of 80% cyanide removal by the RO is based on all cyanide being present as free cyanide, which is unlikely since the CN destruct circuit does not destroy ferrocyanide complexes and weak acid dissociable complexes are also expected to be present. The RO rejection for WAD cyanide will increase in proportion to the difference between the WAD cyanide and free cyanide concentrations. As the RO rejection for cyanide increases, it becomes possible to treat a greater volume of water from the TSF without altering any other assumptions. Table 4-15 shows flows of TSF reclaim at 1.3 mg/L of CN_{WAD} which may be treated in the WTP for different RO cyanide removal efficiencies.

Table 4-15: TSF Treatable Flows at 1.3 mg/L of CN_{WAD} vs. CN_{WAD} RO Rejection

RO Rejection % for Cyanide	Flow of TSF Reclaim to WTP @ 1.3 mg CN_{WAD}/L
80%	7 m^3/h
85%	12 m^3/h
90%	19 m^3/h

In addition, there is also considerable conservatism built in to the assumptions which result in a TSF reclaim water concentration of 1.3 mg CN_{TOT}/L. The key factors which contribute to this uncertainty are the following:

- CN destruct performance – 10 mg CN_{TOT}/L assumed as upper limit for performance. It is realistic to more normally expect operation to be in the range of 0.5 to 5.0 mg CN_{TOT}/L.
- Extent of natural degradation – Based on Figure 4-3 below, greater than 50% natural degradation (attenuation) may be possible with the extended residence times that are typical in tailings ponds. Table 4-16 indicates that, with sufficient time, it is possible to attain a very low concentration of CN_{WAD} with natural degradation alone. At a pH below 9.0, as is expected for the Donlin Gold TSF, WAD cyanide will be expected to dissociate and some of the resultant free cyanide will be lost from the pond surface via volatilization.
- Relationship between total cyanide and WAD cyanide – The limits for CN destruct are based on the total cyanide concentration. The discharge objective is for WAD cyanide. For the purposes of the initial calculations, WAD cyanide has been assumed to be equal to total cyanide. However, as shown in Table 4-16, WAD cyanide will normally be only a fraction of the total cyanide concentration. This is supported by the knowledge that ferrocyanide complexes will not be destroyed in the CN destruct circuit and should, therefore, be expected to represent a portion of the total cyanide concentration in the TSF.

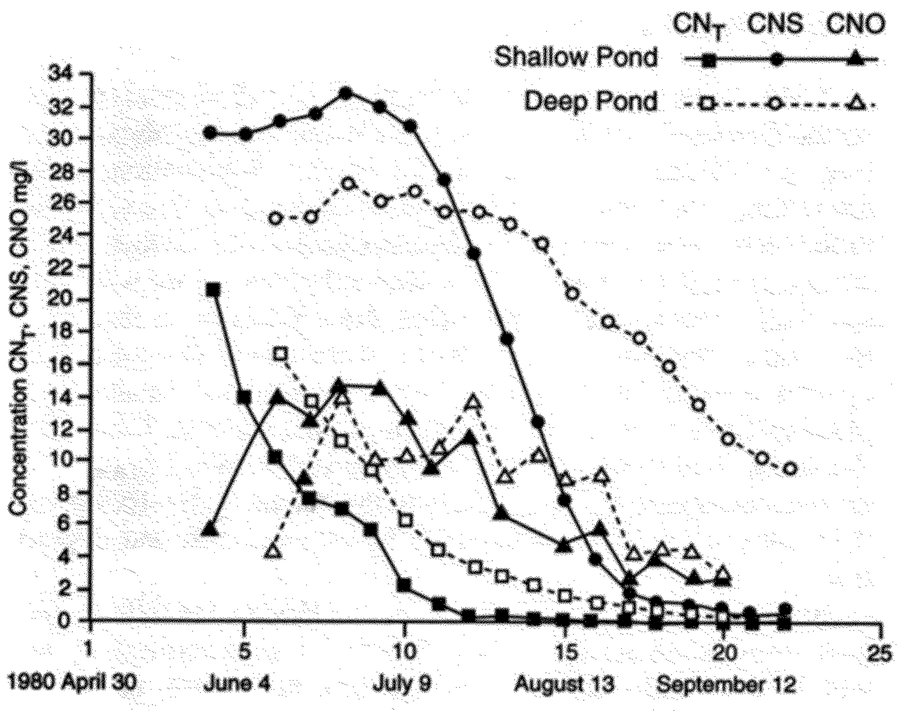


Figure 4-3: Natural Cyanide Degradation in a Northern Canadian Mine [18]

Table 4-16: Effluent Quality of Canadian Gold Mines Applying Batch Natural Cyanide Attenuation Systems [13]

Mine	Location	Barren Bleed mg/L		Final Effluent mg/L	
		Total Cyanide	WAD Cyanide	Total Cyanide	WAD Cyanide
Dome Mines	Porcupine, Ontario	100	98.6 (1983)	0.04	0.02 (1983)
Lupin Mines	Contwoyto, N.W.T.	223	186	0.2	0.02 Sept. 1984)
Cullaton Lake (two ponds)	Keewatin District	800	140 (1982)	--	<0.1 (sept. 1984)

Table 4-17 presents the resulting WAD cyanide concentrations in the TSF reclaim water and treatable flows of TSF reclaim water (based on 80% RO rejection for cyanide) for a few different scenarios to illustrate the influence of the parameters discussed above on the treatment calculations.

Table 4-17: TSF Treatable Flows as Functions of CN Destruct Performance, Natural Degradation and CN_{WAD} / CN_{TOT} Ratio

CN Destruct Performance	Natural Degradation	CN_{WAD} % of CN_{TOT}	TSF Reclaim CN_{WAD} Conc.	Max. TSF Flow to WTP
10 mg CN_{TOT}/L	50%	100%	1.3 mg/L	7 m^3/h
5 mg CN_{TOT}/L	50%	100%	0.66 mg/L	14 m^3/h
1 mg CN_{TOT}/L	50%	100%	0.13 mg/L	85 m^3/h
5 mg CN_{TOT}/L	75%	100%	0.26 mg/L	35 m^3/h
5 mg CN_{TOT}/L	50%	50%	0.33 mg/L	27 m^3/h

In order to evaluate TSF treatability with respect to cyanide concentration, a 1 mg CN_{WAD}/L concentration will be assumed exiting CN destruct, with 50% degradation in the pond, resulting in a reclaim concentration of 0.13 mg CN_{WAD}/L under "summer" conditions and a conservative concentration of 0.26 mg CN_{WAD}/L during winter when natural degradation is minimal [9]. On this basis, it could be possible to blend between 35 and 85 m^3/h into the treatment mix.

Since these flows are higher than the TSF flow limitation imposed by mercury (refer to Section 4.4.4.3), the constraints due to mercury concentration in the TSF are used to establish the average and peak flows for TSF reclaim water to the WTP.

4.4.4.3 Mercury

Treatment of TSF water is limited to 7 m^3/h based on average flows (Dewatering Wells, SRS and Upper CWD) and the following conservative assumptions:

- No removal by the upstream processes (HRC and greensand media filter);
- 94% mercury removal by RO;
- 10 ppb mercury in the TSF reclaim water; and
- Maximum 1.3% TSF seepage in the SRS water (Year 0 – 10).

The 10 ppb mercury concentration in the TSF reclaim represents an assumed worst-case value based on removal efficiencies with the UNR-921 Mercury Control reagent that are being surpassed at other Barrick operations.

Removal efficiency by RO is conservatively taken to be 94%. Removal efficiency of 96-98% at neutral pH has been quoted elsewhere [16] and is generally accepted as possible by the industry.

Seepage fractions in SRS water are expected to increase with time. During the initial ten years of operation, the maximum expected fraction is 1.3%. The maximum LOM fraction is predicted to be up to 10.7%. In addition to the increase with time, there is also a seasonal effect. Maximum seepage rates occur typically in late winter (April). After April and during the entire period when the WTP is generally planned to operate (April – October), the maximum and average seepage fractions are significantly lower. Table 4-18 presents the expected seepage fractions for year 18 and thereafter [17]. Before year 18 (May to October), maximum seepage fractions are lower than 1.3% and for this reason are not shown in the table. Further information on the time influence on the SRS seepage fraction and resulting effect on mercury concentration is presented in Appendix A.

Table 4-18: SRS Seepage Fractions

Year	Seepage Fractions (%)				
	Max (LOM)	Max (Year 0-10)	Max (April)	Avg (May to October)	Max (May to October)
18	10.7	1.3	4.6	1.2	1.3
19			5.2	1.6	2.5
20			8.4	1.7	3.5
21			8.5	2.3	3.8
22			5.2	2.5	5.2
23			7.2	2.1	3.3
24			10.2	2.1	2.6
25			8.4	2.3	2.9

From Table 4-18, the seepage fractions will remain lower than 1.3% until year 18, except during April. Therefore, treatment of at least 7 m^3/h of TSF water will be possible. In contrast, from year 18 onwards, the possible presence of higher percentages of TSF seepage in the SRS water may increase the mercury concentration in the SRS water. In turn, this may require restrictions to be placed on treatment of TSF water and, also, may prevent treatment of the full flow from the SRS during low periods, typically in late winter (refer to Appendix A). If it becomes necessary to limit SRS flow to the WTP in order to respect the discharge objectives for mercury, any excess SRS flow would be used in the process plant and ultimately end up in the TSF.

It should be noted that the predicted SRS seepage is based only on an EPA estimate of liner leakage rates. In addition, there is opportunity to optimize the WTP performance by this point in operation.

Table 4-19 presents treatable TSF flows as a function of UNR reagent and RO removal efficiencies until year 18 of operation:

Table 4-19: TSF Treatable Flows (Until Year 18)

	94% RO Removal	96% RO Removal	98% RO Removal
10 ppb Hg in TSF	7 m^3/h	13 m^3/h	33 m^3/h
5 ppb Hg in TSF	17 m^3/h	30 m^3/h	71 m^3/h

Treatment of TSF reclaim water is constrained by mercury, but treatment is possible at flows in the range of 7-71 m^3/h based on the mercury concentration in the TSF reclaim water, mercury concentration in the SRS water and RO removal efficiency.

It is reasonable to consider treatment of TSF reclaim water at an average design flow of ~30 m^3/h. This flow is associated with a mercury concentration of 5 ppb in the TSF reclaim water and 96% removal at the RO stage. A peak flow of 71 m^3/h has been considered in defining the WTP design capacity.

4.4.4.4 Selenium

Selenium can take a variety of forms; however, most notably in the mining industry, selenium will occur as: (a) selenide (Se^{-2}), which is an impurity in metallic sulphide ores; (b) selenium dioxide, which is oxidized selenide that forms when ores are roasted; (c) selenious acid (H_2SeO_3), which forms when ore roasting flue gas is quenched with water; and (d) wastewater forms of selenium, which are selenite IV (SeO_3^{-2}), selenate VI (SeO_4^{-2}), and solid selenide.

The waste rock at Donlin Gold is the primary source for selenium and, consequently, the Upper CWD water is the main source of selenium in the feed to the WTP. Selenium in the feed to the WTP is expected to be in its highest oxidation state and, thus, will be present as selenate, Se (VI) [6].

No removal of selenate is anticipated in the pre-treatment steps (HRC / greensand filters). Iron co-precipitation is able to remove selenium as selenite (Se IV) but is unable to remove selenate (Se VI) [5].

Assuming 91% selenium removal by RO (see Table 4-14) and blended feed containing 0.038 mg/L selenium, enough removal should be obtained in the RO step to meet the 0.005 mg/L discharge objective.

It is important to consider that RO selenium removal could be insufficient for blends of water containing a contribution of greater than 35% from the Upper CWD water source.

However, it is important to note that the 95th percentile selenium concentration in the Upper CWD during the first five years of operation is expected to be about half of the selenium concentration considered in the design basis.

In addition, the selenium removal efficiency used to evaluate RO performance is possibly conservative. A removal efficiency of 97% has been cited elsewhere [16].

Considering that feed water will be more dilute during the initial operating period (year 0-5) and RO removal efficiency may be higher than 91%, a polishing step to remove selenium from the RO permeate will not be required from the beginning of operations. This initial period should serve to obtain enough field data to evaluate the actual need of a polishing step.

As a polishing step, either conventional IX processes (SBA resins) or BioteQ's proprietary Selen-IX™ process may be employed if necessary. Refer to Section 4.2.2.3.

Selen-IX™ uses an anionic resin in SO_4 form. Since there is reduced SO_4 competition, the process can be used either before or after RO [4].

5. Summary

A conceptual Advanced Water Treatment (AWT) process has been developed for Donlin Gold as an alternative to the High Density Sludge treatment option included in the Feasibility Study. The AWT process, consisting of iron co-precipitation, greensand filtration and reverse osmosis, satisfies the goal to produce treated water meeting the discharge objectives while also allowing treatment of excess water from the TSF, Upper CWD, and SRS. It is capable of treating TSF reclaim water as part of a blended feed along with flows from the dewatering wells, SRS and Upper CWD. This permits the AWT system to play a role in managing and controlling the rate of water accumulation in the TSF and provides a tool to enhance site water management and operational flexibility.

6. References

[1] American Water Works Association. Treatment Process Selection for Particle Removal. Page 77, Figure 4.1.

[2] Aube, B. and Stroiazzo, J. 2000. Molybdenum Treatment at Brenda Mines. Proceedings of the 5th International Conference on Acid Rock Drainage.

[3] BGC. 2014. "Water Resources Management Plan Interim Update", Rev. 1. (Report No. PM-0011178-0040).

[4] BioteQ Environmental Technologies. 2014. Selenium Removal from Mine Impacted Waters Using Selen-IX™.

[5] CH2M Hill. 2010. Review of Available Technologies for the Removal of Selenium from Water. North American Metals Council.

[6] Day, S. 2015a. E-mail "RE: CWD Selenium oxidation state Question". Received February 19, 2015.

[7] Day, S. 2015b. Revised Process Pond Water Predictions – DRAFT. SRK. April 15, 2015.

[8] Degremont. Memento Technique de l'Eau – Degremont. 10th edition.

[9] Fueg, J. 2015. E-mail "RE: Treatment of TSF Water – Cyanide". Received April 13, 2015.

[10] Guo, X et al. 2009. Removal of Antimony (V) and Antimony (III) from Drinking Water by Coagulation – Flocculation - Sedimentation (CFS). Water Research 43 (2009) 4327-4335.

[11] Hernandez, A. 2015. Preliminary Design Basis, Rev. 0. (H348428-00000-210-226-0001).

[12] Krumins, T. 2015. Process Plant Water Management Philosophy, Rev. 0. (H348428-00000-210-226-0002).

[13] Meech, J. "Cyanide Destruction Methods MINE 292 – Lecture 19". http://www.jmeech.mining.ubc.ca/MINE290/MINE292-Lecture19-Cyanide%20Destruction-2013.pptx.

[14] Ministère du Développement durable, de l'Environnement et des Parcs du Québec (MDDEP). 2006. Guide de conception des installations d'eau potable – Volume 1.

[15] Mining Association of Canada. 2008. Application of Membrane Separation Technology to Mitigation of Mine Effluent and Acid Drainage. MEND Report 3.15.1.

[16] Pure Water Gazette. 2012. Reverse Osmosis Membranes – Typical Rejection Characteristics. http://www.purewatergazette.net/blog/reverse-osmosis-membranes-typical-rejection-characteristics/

[17] Rieser, M. 2015a. E-mail "RE: SRS Geochem 50-95-Max WQ estimates - Updated April 15-2015.xlsx". Received April 21, 2015. File "TSF underdrain mercury mass loading.xlsx".

[18] Schmidt, J. W. et al. 1981. "Development Studies for Suitable Technologies to Remove Cyanide and Heavy Metals from Gold Milling Effluents". Proc. 36th Industrial Waste Conf., Purdue University, Lafayette, Indiana, p. 831-849.

[19] U.S. Environmental Protection Agency. 2014. Reference Guide to Treatment Technologies for Mining-Influenced Water.

[20] Veolia Water Systems. Mining Brochure. http://www.veoliawaterst.com.au/vwst-australia/ressources/documents/2/8692,VWS_Australia_Mining-Brochure_2011.pdf.

Appendix A: SRS Water Quality – Influence of Mercury Concentration on Treatment Flows from SRS and TSF Water Sources

SRS source water quality is influenced by the TSF seepage flow. The TSF seepage fraction is expected to increase with time. Figure A1 presents the SRS source water fractions during the Life of Mine. Maximum LOM seepage fraction has been estimated to be 10.7%.

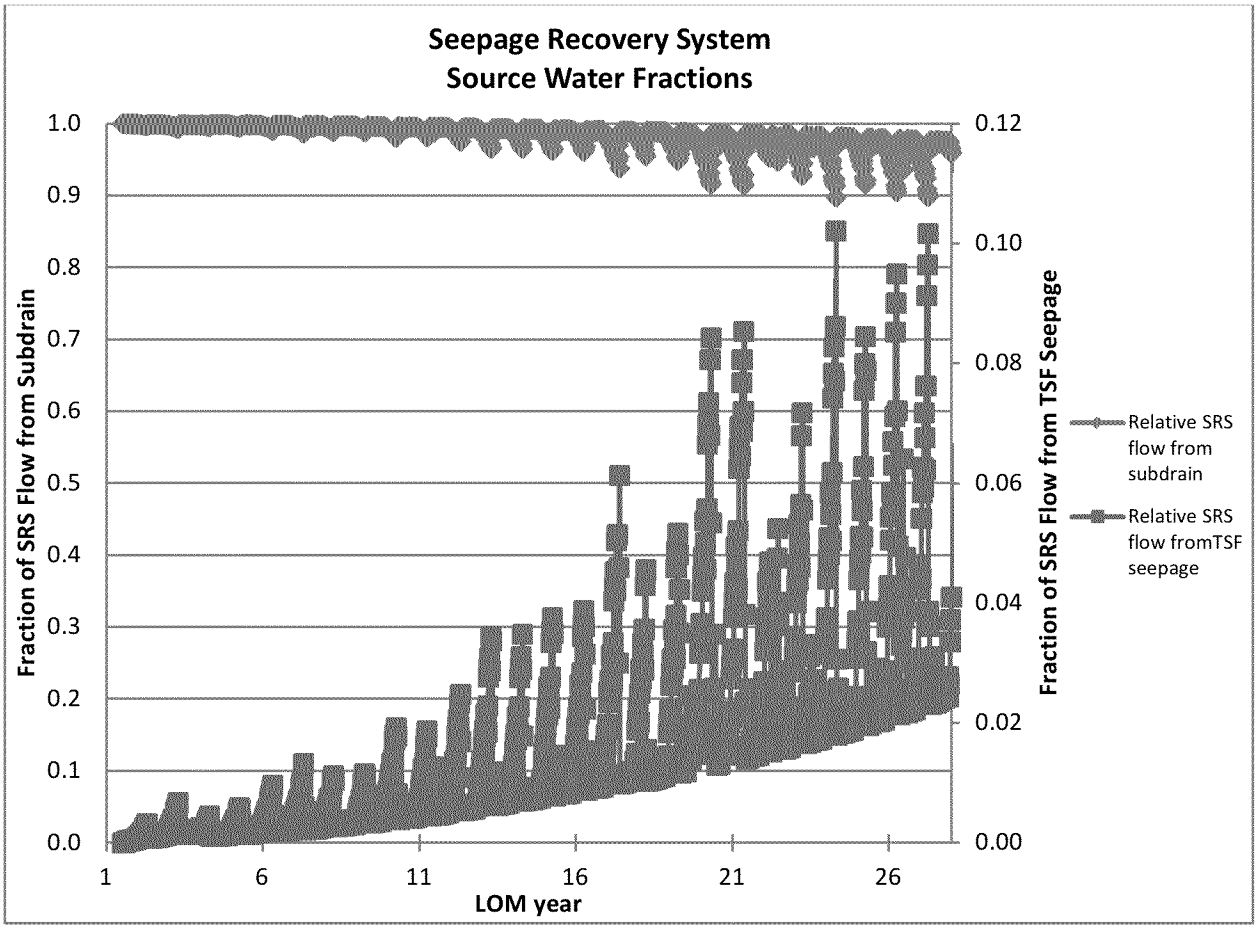


Figure A1: Seepage Recovery System – Source Water Fractions

High seepage fraction may have an impact on treatability of SRS water, the limiting factor being mercury concentration. Figure A2 shows mercury concentration in the SRS water as a function of the fraction of SRS flow from TSF seepage. Mercury concentration has been estimated assuming 10 ppb in the TSF and 95th percentiles for the other water sources. Maximum mercury concentration in the SRS water has been estimated at 0.00108 mg/L.

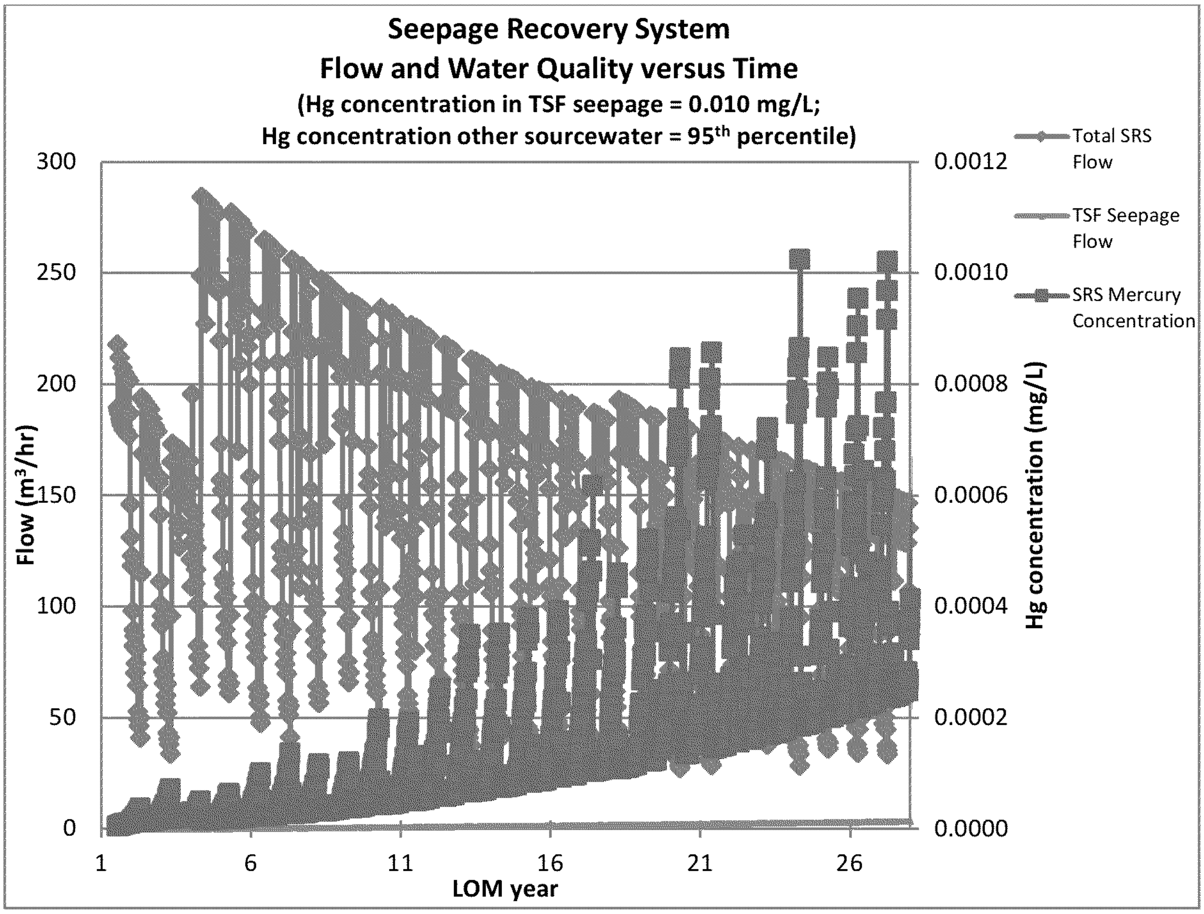


Figure A2: SRS – Flow and Water Quality Versus Time

As can be seen from Figure A2, the mass loading increases at a relatively constant rate over the LOM as the TSF volume and seepage increase and the clean water entering the underdrain decreases. However, on a seasonal basis, the concentrations increase dramatically during the winter as the clean water entering the underdrain decreases. Since the water treatment plant will be operated during the summer period, the effect of seasonal differences in water quality needs to be considered.

In addition to the impact on SRS treatability, the TSF seepage fraction may also affect TSF water treatability. Table A1 presents treatable SRS and TSF flows versus RO removal efficiency for mercury and mercury concentration in the TSF for Years 0 - 17. The maximum 95th percentile SRS mercury concentration during Years 0-17 (May to October) is 0.000151 mg/L. This concentration has been used for calculation.

Table A1: SRS and TSF Flows – Years 0 – 17 (May to October)

RO efficiency removal (%)	SRS flow treated (m^3/h)	TSF flow treated (m^3/h)	
		10 ppb in TSF	5 ppb in TSF
94	165	7	17
96	165	13	30
97	165	20	44
98	165	33	72

From Table A1, it can be seen that during operation between May and October in Years 0 – 17 it would be possible to treat all SRS flow and between 7 – 72 m^3/h of TSF reclaim water, depending on actual RO removal efficiency and mercury concentration in the TSF. It would be reasonable to expect a TSF reclaim water flow to treatment between 20 – 30 m^3/h, which means RO removal in the range 96 – 97% with a TSF mercury concentration of between 5 – 10 ppb. It is important to note that some mercury concentration peaks may happen in April after Year 13, resulting in mercury concentrations in the SRS water up to 0.00037 mg/L. As shown in Table 2 below, these higher mercury concentrations will cause no impact on SRS flow. Conversely, a significant decrease in TSF flow may happen.

Table A2 shows the impact of TSF seepage fraction on SRS and TSF water sources treatability for Year 18 and later.

Table A2: SRS and TSF Flows – Year 18 and Later (April – October)

Year	SRS Seepage Ratio			SRS Hg Concentration (mg/l)			SRS Treated Flow (m^3/h)			TSF Treated Flow (m^3/h)		
	Max (April)	Avg (May to October)	Max (May to October)	Max (April)	Avg (May to October)	Max (May to October)	Min (April)	Avg (May to October)	Min (May to October)	Min (April)	Avg (May to October)	Min (May to October)
18	4.6	1.2	1.3	0.000465	0.000125	0.000135	165	165	165	2.5	7.5	7
19	5.2	1.6	2.5	0.000525	0.000165	0.000255	165	165	165	2	6.5	5.5
20	8.4	1.7	3.5	0.000845	0.000175	0.000355	115	165	165	0	6.5	4.5
21	8.5	2.3	3.8	0.000855	0.000235	0.000385	110	165	165	0	6	3.5
22	5.2	2.5	5.2	0.000525	0.000255	0.000525	165	165	165	2	5.5	2
23	7.2	2.1	3.3	0.000725	0.000215	0.000335	135	165	165	0	6	4.5
24	10.2	2.1	2.6	0.001075	0.000215	0.000265	80	165	165	0	6	5.5
25	8.4	2.3	2.9	0.000845	0.000235	0.000295	115	165	165	0	6	5.5

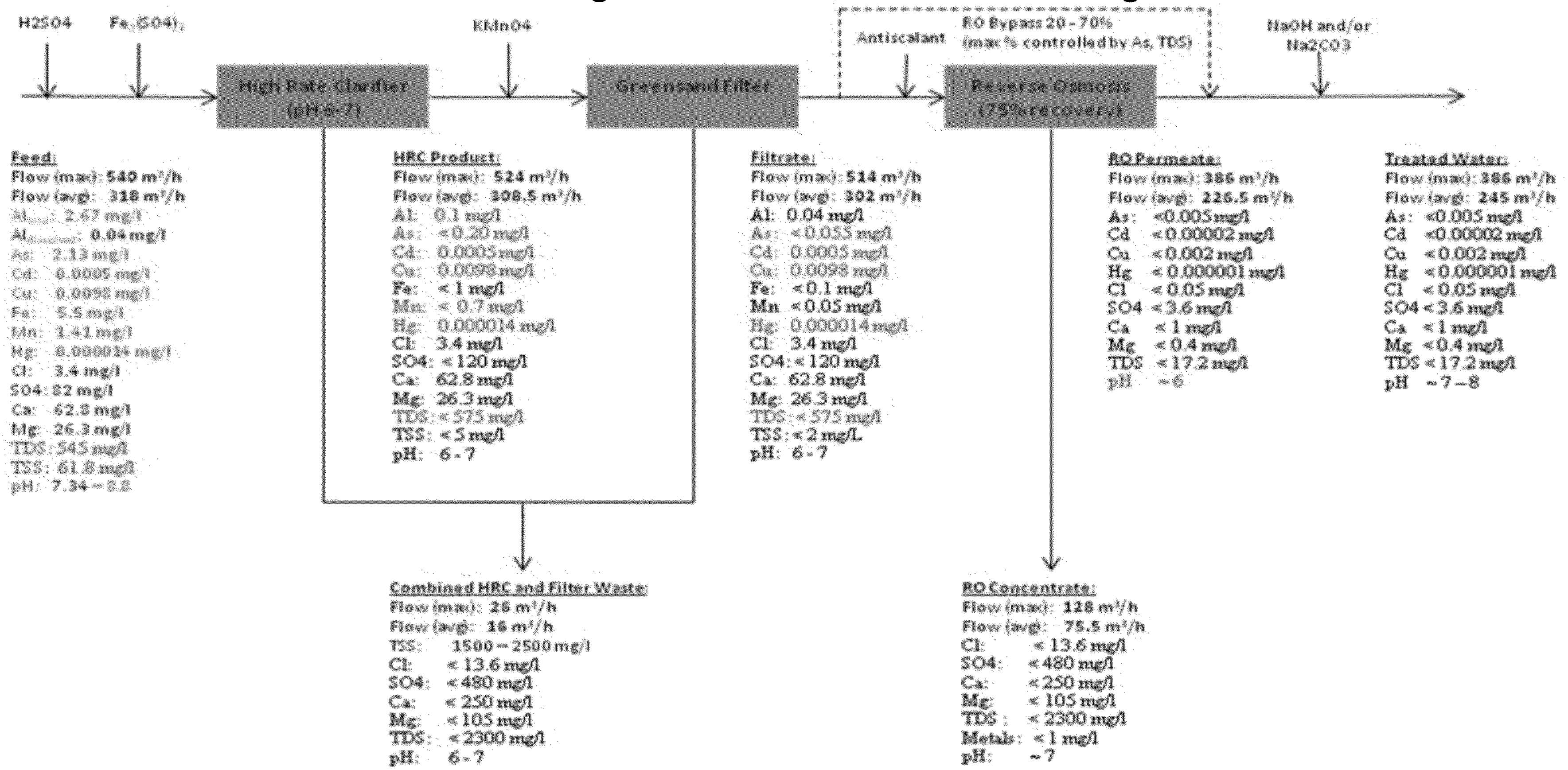
Notes: Hg concentration in TSF 10 ppb
Mercury RO removal efficiency 94%
Values highlighted in yellow indicate treatment restriction

From table A2, it can be seen that the first appearance of high mercury concentrations in the SRS water causing an impact on treatability is expected to happen in April of Year 20. From that year, high seepage ratios and consequently high mercury SRS concentration will happen during winter time, mainly in April (maximum ratio 10.2% in April, Year 24). An exception is Year 22, where the maximum occurs in early June. During the rest of the operation period (May to October), average ratios in the range of 1.2 – 2.5% can be expected.

From this analysis, it can be concluded that during the operation period comprising Years 18 – 25 of the life of mine, assuming conservative values of 10 ppb mercury concentration in the TSF reclaim water and 94% RO mercury removal efficiency, some decrease in the treatable SRS flow and minimal TSF treatment will occur in April. During the rest of time (May – October), no SRS flow restriction and only minor decrease of TSF flow are expected.

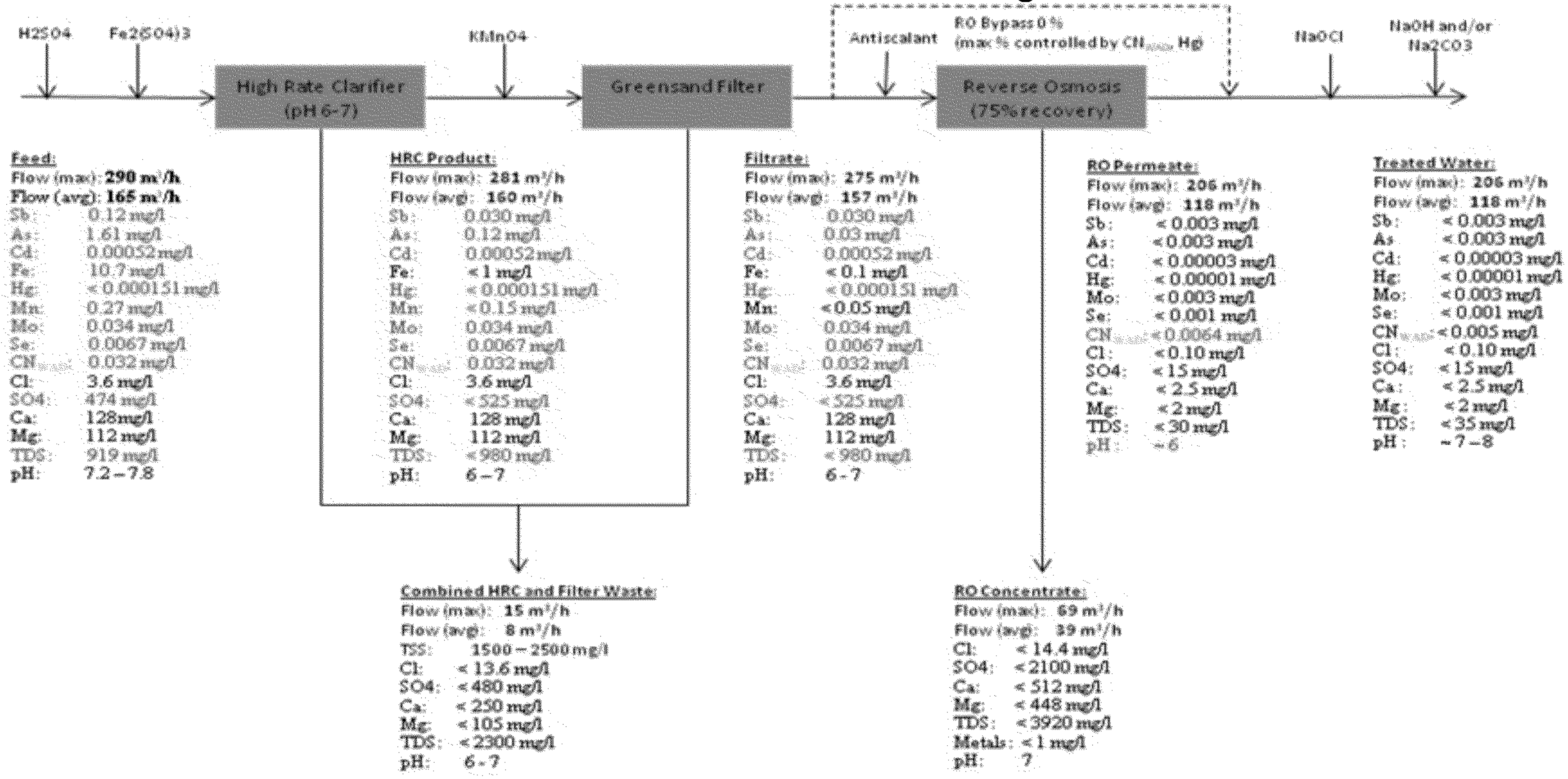
Appendix B: Treatment Flowsheets for Individual Water Sources

Dewatering Wells Water Treatment Flow Diagram



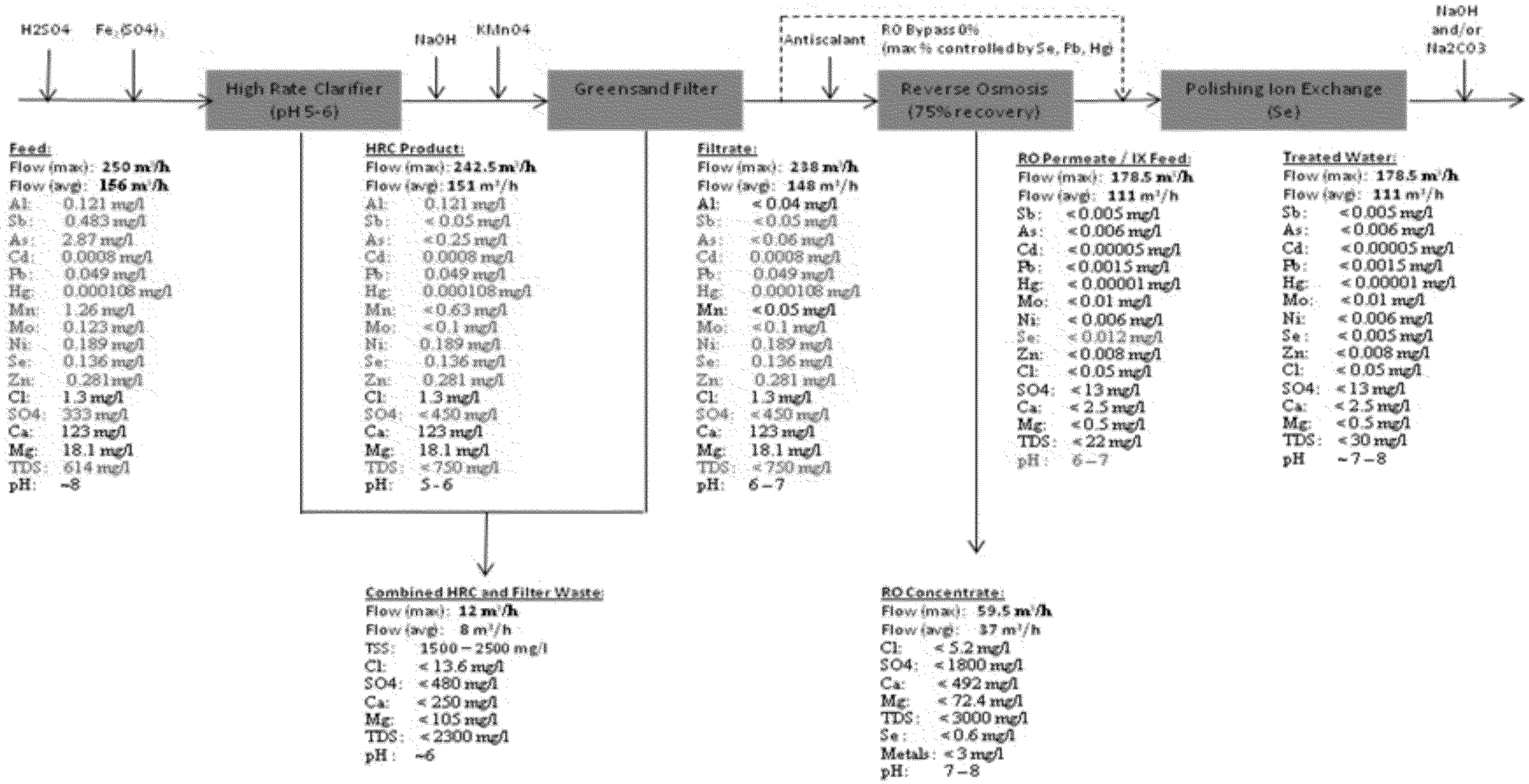
Notes: 1.- Feed concentrations are the 95th percentile concentrations used for design
2.- Average flows based on summer treatment (April to October) at 100% availability
3.- Peak flows are peak weekly averaged flows from BGC water balance
4.- For calculation purposes RO by-pass has not been taken into account
5.- The High Rate Clarifier (HRC) can be operated at a pH of 6-7 when treating only Well water due to low feed Sb

SRS Water Treatment Flow Diagram



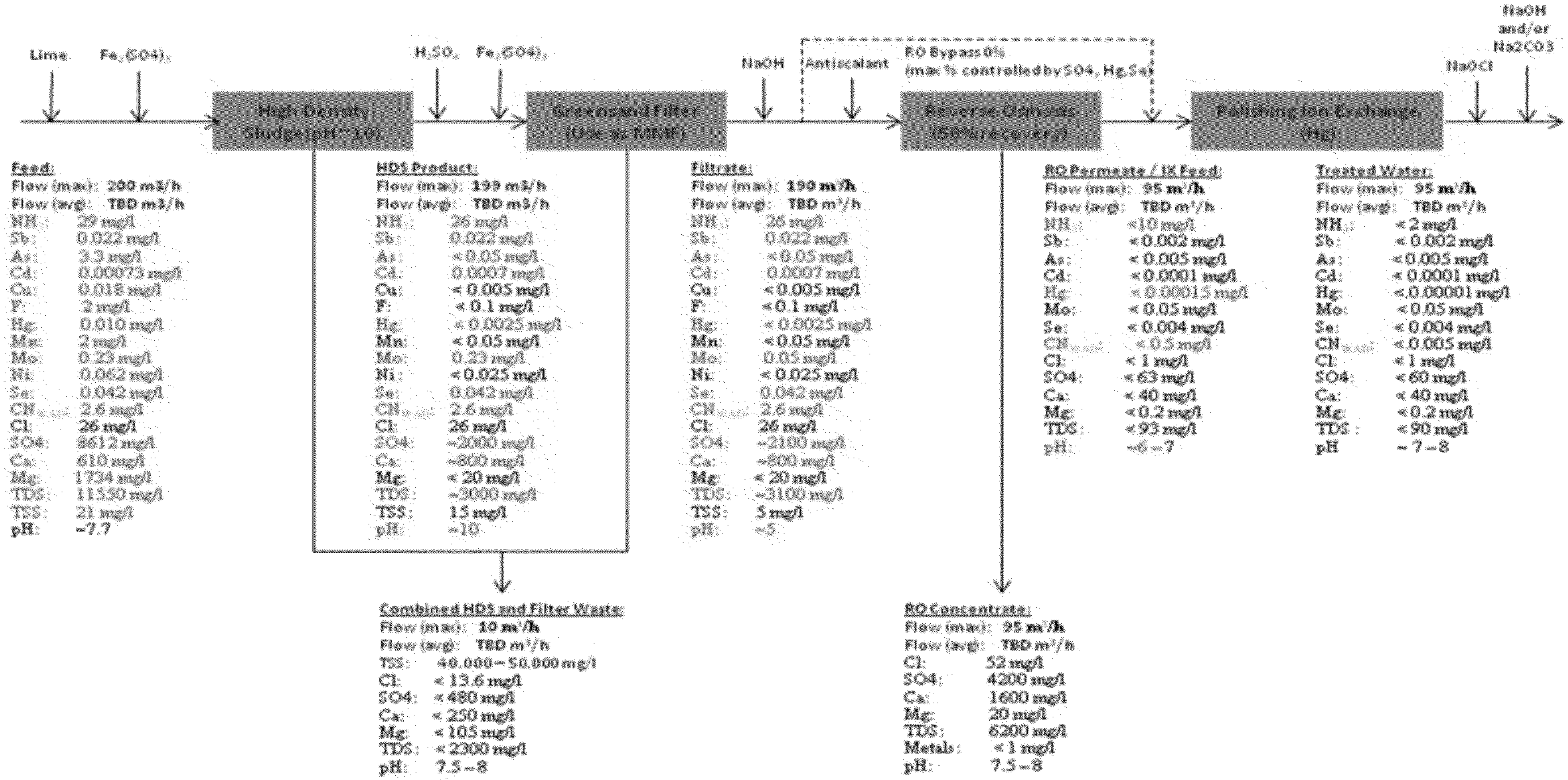
Notes:
1.- Feed concentrations are the 95th percentile concentrations used for design
2.- Mercury feed concentration is 95th percentile value prior to Year 18. Higher concentrations may result after Year 18 in late winter with low SRS underdrain flow
3.- Average flows based on summer treatment (April to October) at 100% availability
4.- Peak flows are peak weekly averaged flows from BGC water balance
5.- Operation pH during initial operation period (first 17 years): 6-7. Ferric sulfate dosage may need to be increased with lower operating pH (5-6) afterwards to cope with higher Sb concentration
6.- Treated water result for WAD cyanide is based on 80% RO removal efficiency (most conservative assumption). Lower WAD cyanide concentrations in the treated water may be attainable by improved RO rejection without need for oxidation of residual WAD cyanide using NaOCl

Upper CWD Water Treatment Flow Diagram



Notes: 1.- Feed concentrations are the 95th percentile concentrations used for design
2.- Average flows based on summer treatment (April to October) at 100% availability
3.- Ferric sulfate dosage has been based on achieving an Fe concentration of 6 x 10-4 mol/l in the HRC. Fe dose may need to be increased for greater Sb removal
4.- Selenium removal performance is expected to constrain the ability to treat CWD water alone unless a polishing IX step is added to achieve compliance with the discharge objectives

TSF Water Treatment Flow Diagram



Notes: 1.- Feed concentrations are the steady-state concentrations used for design

2.- The feed WAD cyanide concentration reflects winter operation with no natural degradation assumed to take place in the TSF. It is assumed that final oxidation of WAD cyanide and ammonia can be performed with NaOCl to meet the discharge objectives. This would require further validation/testing

3.- Improved molybdenum removal can be achieved, if necessary, by reducing pH and adding ferric sulfate before the MMF to take Mo out via co-precipitation / adsorption mechanisms

4.- A maximum flow of 200 m3/h has been assumed for individual treatment of TSF reclaim water for the purposes of illustrating the flow balance. This is roughly aligned with discharge of 800,000 m3 per year for operation 350 days/year.